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CONTROL OF ATMOSPHERIC EMISSIONS IN THE WOOD PULPING INDUSTRY. VOLUME 2

E. R. Hendrickson, et al

Environmental Engineering, Incorporated
Gainesville, Florida

15 March 1970

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DIVISION OF ECONOMIC EFFECTS RESEARCH

CONTROL OF ATMOSPHERIC EMISSIONS IN THE WOOD PULPING INDUSTRY

FINAL REPORT
CONTRACT NO. CPA 22-69-18
MARCH 15, 1970

VOLUME 2

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CHAPTER 5

CONTROL METHODS PRESENTLY IN USE

SUMMARY

Control methods presently in use in the wood pulping industry consist of add-on hardware or process modifications. The methods considered in this chapter are those which have been in reasonably successful operation for at least one year at one or more locations. The one exception is the recently developed recovery systems which eliminate direct contact between the flue gases and black liquor. Control methods are briefly described in general terms and are evaluated under conditions of specific applications. The evaluations include a cost and effectiveness study as well as a discussion of engineering factors which are unique to the application. Cost calculations have been prepared for the total capital cost and net annual cost. It must be recognized that installations vary widely from mill to mill, thus the costs presented can serve only as a general guide. Cost estimates for specific mills must be based on the situation at hand.

A variety of methods is found to be useful in controlling particulate emissions from pulp mill sources. Efficiencies of 99 + percent are possible. The listing which follows identified those methods most commonly used and which are described and evaluated in the chapter. It should be noted that the direct contact evaporator following the kraft recovery furnace is an important particulate control device. It should also be noted that where primary or secondary wet scrubbers are used as particulate collectors on combustion sources, the emission of gaseous sulfur compounds may be increased or decreased depending on the nature of the scrubbing medium.

Kraft Recovery Furnace	- Electrostatic Precipitators - Venturi Evaporator/Scrubbers - Electrostatic Precipitators plus Secondary Scrubber
Kraft Lime Kiln	- Venturi Scrubber - Cyclonic Scrubber - Impingement Baffle Scrubber

Kraft Smelt Dissolving Tank	- Mesh Pads - Packed Tower Scrubbers - Orifice Scrubbers
Kraft Lime Slaker	- Mesh Pads - Cyclonic Scrubbers
Power Boilers	- Mechanical Collectors - Electrostatic Precipitators
Combination Boilers	- Mechanical Collectors - Cyclonic Scrubbers

Fewer numbers of methods are in use for the control of gaseous emissions, particularly the odorous reduced sulfur compounds. The majority of odorous emissions can be grouped into three categories: (a) recovery furnace offgases; (b) low volume high concentration sources such as the noncondensible gases from the multiple effect evaporators, and digester relief and blow gases; (c) high volume low concentration sources such as brown stock washers and smelt dissolving tank. Wet scrubbers have received limited application on combustion sources in the kraft process because of the difficulty of absorbing effectively all of the odorous compounds with a single scrubbing medium. Also as indicated previously, the emission of gaseous sulfur compounds may be increased or decreased depending on the nature of the scrubbing medium. The following list identifies those methods most commonly used and which are described and evaluated in the chapter:

Kraft Recovery Furnace	- Weak Black Liquor Oxidation - Strong Black Liquor Oxidation - Proper Operation - Venturi Evaporator/Scrubber - Cyclone Evaporator/Scrubber - New Recovery System Design
Kraft Smelt Dissolving Tank	- Packed Tower Scrubber - Orifice Scrubber
Kraft Digester Relief and Blow plus M. E. Evaporator	- Chlorination - Incineration - Packed Tower Scrubbers - Weak Black Liquor Oxidation (M.E. Evaporators)
Sulfite Acid Tower	- Additional Absorption Tower
Sulfite Blow Pit	- Packed Tower

5.1 INTRODUCTION

Control methods presently in use in the wood pulping industry consist of add-on hardware or process modifications. The methods considered in this report are those which have been in reasonably succesful operation for at least one year at one or more locations. The one exception to this is the inclusion of recently developed recovery systems which eliminate direct contact between the flue gases and black liquor. Several of these systems are being installed and are described because of the potential they offer for reducing odorous emissions and the interest expressed in these units by the industry. For a summary of the estimated number of control methods presently in use in the industry, the reader is referred to Chapter 12, Table 12-1.

In this chapter, control methods are briefly described in general terms and are evaluated under conditions of specific applications. The evaluations include a cost and effectiveness study and a discussion of engineering factors which are unique to the application.

Cost calculations are prepared for the total capital cost and the net annual cost. These calculations and the assumptions and definitions used are included in Section 5.3. It should be emphasized that the costs presented in this report are based on generalized unit cost data and certain specified assumptions. The costs can be very useful for comparing several control methods as applied to a specific source, but they should not be used as an absolute cost estimate for the application of a specific control method to a specific source. The application of control methods from mill to mill varies considerably depending upon such factors as the regional cost index, the physical arrangement of the equipment, and the capacity of equipment. For this reason cost estimates for specific installations must be based upon costs and other considerations unique to the situation at hand.

The capital costs included in this report are based on January, 1969, prices. Since it would be desirable to be able to escalate these capital costs to meet future conditions an attempt was made to prepare a curve for escalation of these costs. Due to the diverse nature of the control equipment and process changes, it was found that there was not one set of curves that could be used to escalate the capital cost of all control methods on an individual basis. Generalizing then, the contractors believe that the cost for a number of control methods applied to a number of different mills can be escalated by the use of equipment indices prepared by Marshall & Stevens Publication Company, 1717 Beverly Boulevard, Los Angeles, California 90026. These indices are entitled "Equipment-National Average" and are listed under the subheading entitled "Paper Manufacturing." For a more precise capital cost, the cost of individual control methods must be individually recalculated in the future.

5.2 GENERAL DESCRIPTION OF CONTROL EQUIPMENT

5.2.1 ELECTROSTATIC PRECIPITATORS

An electrostatic precipitator consists in principle of a number of discharge (emitting) electrodes, collecting electrode plates and a high-voltage power unit. This power unit comprises high-voltage transformers and rectifiers to convert the available AC power to high-voltage DC power. Lately, silicon diode rectifiers have almost exclusively replaced other types of rectifiers.

The dust laden gas enters the electrostatic precipitator and flows in the passages created by the collecting electrodes. When high-voltage power is applied to the discharge electrodes located in the passages, ionization or corona discharge occurs near the surface of the discharge electrodes. The negative ions attach themselves to dust particles near the electrodes giving the particles a negative electric charge. The charged dust particles are repelled by the discharge electrodes and attracted to the collecting surfaces connected to ground. Here the dust particles lose their electrical charge and are deposited on the collecting electrodes.

The collecting electrodes are provided with a rapping mechanism for dislodging the dust precipitated on the electrodes. This rapping system must be designed very carefully to avoid creation of a dust cloud in the space between the electrodes. Incorrect rapping operation or programming may cause the dust to re-entrain which, in turn, would greatly decrease the dust collecting efficiency of the precipitator. Incorrect rapping is the most common cause for so-called "snow-outs." The rapping mechanism, which is generally operated on an automatic time cycle basis, is either of vibrator or motor driven fall-hammer type. Vibrators are preferred for pulp mill applications rather than the hammer rappers which are used more frequently, for example, on fly ash precipitators.

The high voltage power level is controlled by an automatic control unit. The discharge electrodes are connected to the negative rather than to the positive side of the power supply. A higher voltage can thereby be maintained without excessive arc-overs which, in turn, would cause waste of electric power and eventually lead to operating difficulties. The automatic voltage rate of sparking provides a feedback to the control unit making it possible automatically to operate the electric system of the precipitator at optimum conditions. Little or no sparking will increase the voltage and too much sparking will reduce the voltage, thereby preventing arc-overs detrimental to the precipitator.

Gas conditions are never uniform throughout the precipitator. Aside from unintended irregularities in gas distribution and other conditions, the dust loading varies from inlet to outlet of the precipitator. To reach optimum dust collecting efficiencies, the electrical control units must accommodate these variations. A high dust loading will increase the sparking and reduce the voltage. If the precipitator would have only one electric system with one automatic voltage control, the voltage throughout the precipitator would be limited to the lowest voltage permissible at any point in the precipitator. The modern precipitator is, therefore, divided into independent electrical units, each controlled for maximum voltage depending on the gas conditions in that particular section of the precipitator. The discharge system close to the outlet of the precipitator will consequently operate at a higher voltage level than the inlet. This need for sectionalizing has led to increased use of the horizontal flow, plate-type precipitator. This type has superseded the tube-type, vertical flow precipitator, once very common in the pulp and paper industry. The tube-type precipitator does not easily lend itself to sectionalizing.

The electric power consumption of a precipitator depends on several factors. The input of current is necessary to sustain the voltage level. There is a constant power drain due to ionization in the corona, due to the controlled sparking and other voltage leakages because of poor maintenance, dust build-up, et cetera. The collection efficiency depends on the voltage level; but, on the other hand, the higher the voltage level, the greater the power drain. At 90 percent collection efficiency, the power consumption is approximately 0.2 kw/1000 CFM of gas and at 99.9 percent efficiency, the power consumption has risen to approximately 0.8 kw/1000 CFM.

Most precipitators used in the wood pulping industry on recovery boilers are designed for collection efficiencies of 90 - 99.9 oercent. The pressure drop of the gas passing through the precipitator is usually below 0.5 inch W.G. A typical wet-bottom precipitator with tile shell is shown in Figure 5-1.

5.2.2 VENTURI SCRUBBERS

The Venturi scrubber consists in principle of a convergent section (throat) and a divergent section. Dust laden gas enters the convergent section and is accelerated to high velocity as it approaches the throat. Gas velocities in the throat section vary from 100 to 500 FPS.

Water or other scrubbing liquid is injected either directly into the throat section or the top of the Venturi. In the latter case, the scrubbing liquid cascades down the walls of the convergent section. The high velocity gas stream atomized the liquid into a fine mist--the greater the velocity, the finer the droplets. Collision between the dust particles and the water droplets takes place and causes the dust to be entrapped in the water. Further collision between the water droplets occurs. This will create aggregate droplets of relatively large size. These droplets are easily separated from the gas stream in a subsequent separator. The collision or impaction phenomenon is rather complex, but is mainly due to mass forces created by the great velocity differential between the dust particles and the water droplets. For submicron particles, the Brownian molecular movement, diffusion, and electrostatic forces also play an important role.

In the divergent section, the gas and the dust particles are decelerated thereby creating a new velocity differential with additional agglomeration. Finally, in the elbow connecting the Venturi and the separator, as well as in the inlet to the separator (if radial), changes of direction of the gas flow cause additional impaction and agglomeration.

The gas and liquid enter the separator, usually of the cyclonic type, where the liquid is thrown to the walls by centrifugal forces and drains to the bottom by gravity. The clean gas exits through the upper portion of the separator.

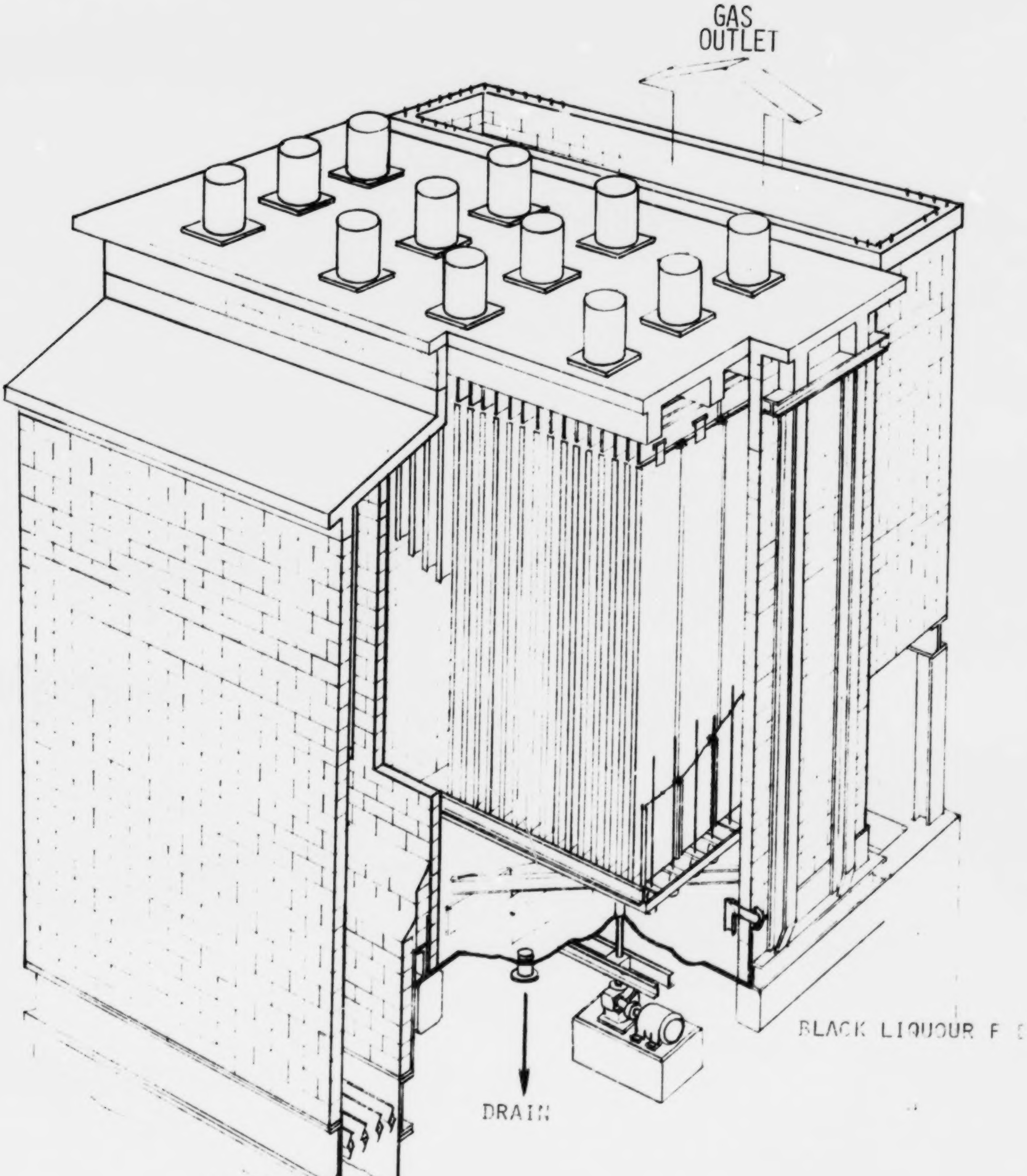

GAS OUTLET
BLACK LIQUOUR F
DRAIN
GAS INLET

FIGURE 5-1

PRECIPITATOR FOR RECOVERY BOILER

The elbow connecting the Venturi with the separator is either of the regular type or of so-called flooded type. The flooded elbow is developed to prevent erosion. Here the liquid surface absorbs the brunt of the water and dust stream. For pulp mill applications, such as lime kiln scrubbers, the flooded elbow is recommended.

Normally the separator is of the cyclonic type with conical bottom and either top outlet or side outlet. The type of outlet depends on whether the fan is located before or after the scrubber. The top outlet lends itself to a direct mounted stack. In order to facilitate the installation, the separator is sometimes provided with a flat bottom. This may sometimes create drainage problems. For special purposes a so-called false bottom is also installed. The false bottom creates an intermediate reserve for the scrubbing liquid where de-aeration can take place making the scrubbing liquid (e.g., black liquor) more suitable to pump. Scrubbers for recovery boilers are almost always provided with facilities for wall wash of the separator. This prevents build-up of black liquor solids which can be a fire hazard, in addition to being a maintenance problem.

The Venturi scrubber is capable of high efficiency collection of dust and fumes even in the submicron range. The efficiency of collection is a function of the pressure drop across the scrubber which in turn is a function of the gas velocity in the throat and the liquid flow rate (liquid to gas ratio). The higher the gas velocity or the liquid flow rate, the greater will be the pressure drop and consequently the greater the efficiency. There is, however, a cut-off point where increased liquid flow rate will have an adverse effect on the collecting efficiency. The throat becomes "flooded." This occurs at about 20 gallons per 1000 CFM. In order to attain high collecting efficiency, the water droplet size distribution has to be in a certain relation to the dust particle size distribution. Too large droplets would reduce the probability of collection drastically, and even if collection would take place the chance for the dust particles to be entrapped in water would also be reduced.

Increased gas velocity (pressure drop) increases the atomization. A fine dust will, therefore, require a higher pressure drop than a coarse dust for the same collecting efficiency.

Scrubbing liquid is injected into the Venturi at low pressure through relatively large jets or open weirs. In order to be able to recirculate the scrubbing liquid, in other words maintain a high concentration of solids, the manufacturers try to completely do away with narrow constrictions and weirs. The liquid rate is normally 3 - 15 gallons per 1000 CFM of gas. A typical Venturi scrubber is shown in Figure 5-2.

5.2.3 CYCLONIC SCRUBBERS

The cyclonic scrubber is an efficient device for removing dust particles two microns and larger and is also a relatively good gas absorber. In this unit, the dust laden gas enters tangentially at the bottom of a cylindrical tower and spirals upward through the scrubber in a continuously rotating path. A spray manifold is located axially in the center of the scrubber with banks of spray nozzles directed radially toward the walls. The spray sweeps across the path of the gas stream intercepting and entrapping the dust particles. The centrifugal motion of the spray imparted by the rotating gas causes the droplets to impinge against the walls of the scrubber and drain to the bottom due to gravitational forces.

In another type of cyclonic scrubber, the spray nozzles are mounted on the wall rather than in a central spray manifold. The advantage with this type is that the nozzles may easily be serviced or replaced while the scrubber is in operation.

The mechanism of collecting particles in the cyclonic scrubber is in essence the same as in a Venturi scrubber; namely, impaction and agglomeration of liquid droplets and dust particles with subsequent centrifugal and gravitational separation.

The liquid pressure ranges from 50 to 400 PSIG. The flow rate is normally 3 - 8 gallons per 1000 CFM.

The cyclonic scrubber is most efficient on relatively coarse dust and the efficiency drops off markedly for particles under two microns. The pressure drop is considerably less than for a Venturi scrubber and ranges normally from 0.5 to 3 inches WG. A typical cyclonic scrubber is shown in Figure 5-3.

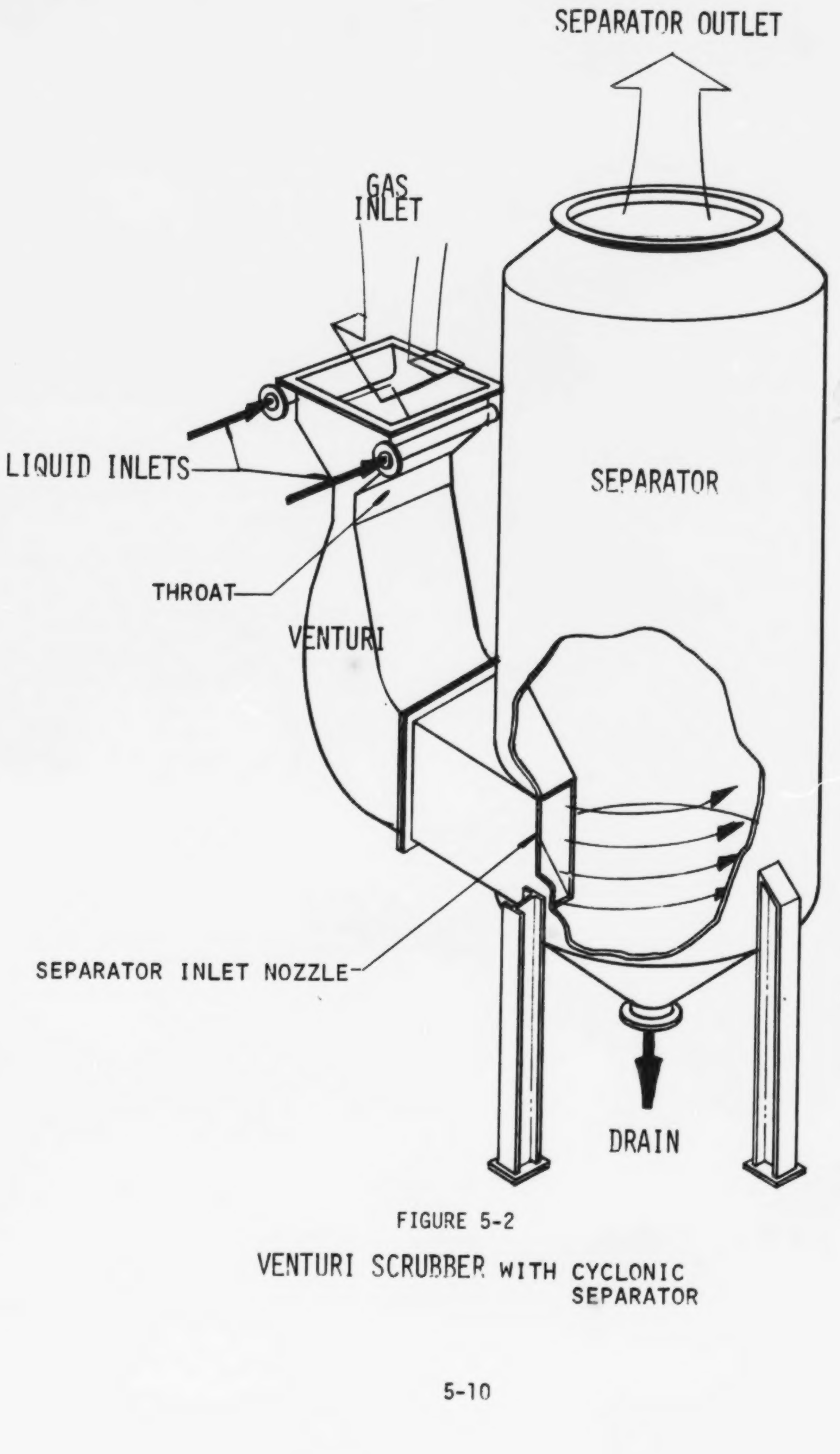

SEPARATOR OUTLET
GAS INLET
LIQUID INLETS
SEPARATOR
THROAT
VENTURI
SEPARATOR INLET NOZZLE
DRAIN

FIGURE 5-2

VENTURI SCRUBBER WITH CYCLONIC SEPARATOR

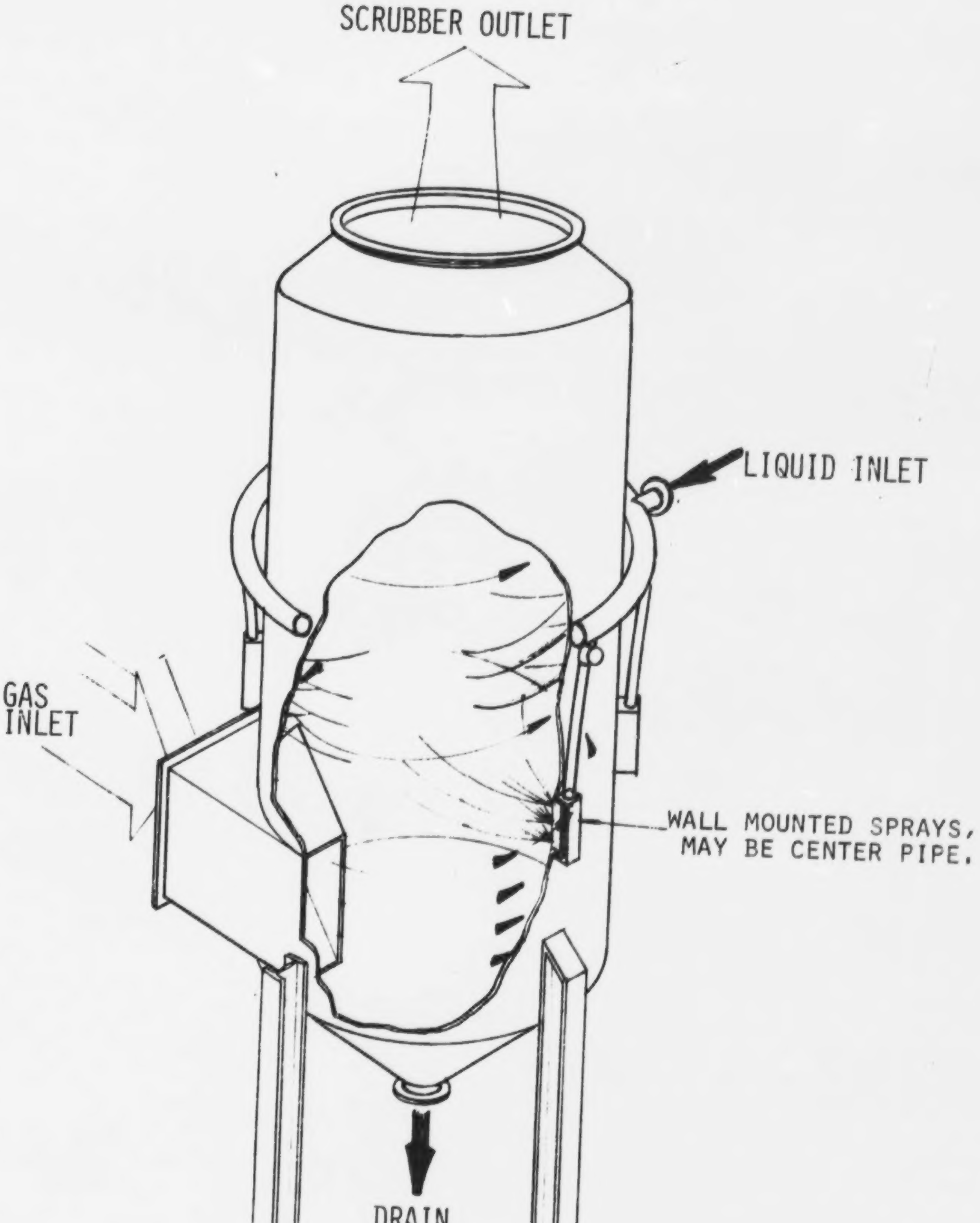

SCRUBBER OUTLET
LIQUID INLET
GAS INLET
WALL MOUNTED SPRAYS, MAY BE CENTER PIPE.
DRAIN

FIGURE 5-3

CYCLONIC SCRUBBER

5.2.4 IMPINGEMENT BAFFLE SCRUBBERS

The impingement baffle scrubber is a vertical tower equipped with one or more impingement baffle stages. The impingement baffle consists of a perforated plate having a multitude of small holes so arranged that a baffle is located directly above each perforation. A weir on each plate maintains a level of scrubbing liquid.

The contaminated gas enters radially at the bottom of the scrubber and is subjected to a water spray that will precipitate out the coarser dust particles. The gas then passes through the perforations and impinges on the baffles. The gas velocity through the perforations, ranging from 75 to 100 FPS, creates an atomization of the scrubbing liquid and the velocity differential between gas and liquid results in the impaction and agglomeration of liquid and particulate matter. The scrubbing liquid is introduced into the scrubber through low pressure spray nozzles located below the plates and spraying upward. These sprays not only deliver liquid to the plates, but also serve to cool, humidify, and condition the gases; remove coarse particles; and keep the bottom of the plates clean. For scrubbers with more than one stage, the scrubbing liquid is drained off the plates down ward from stage to stage in the scrubber. In addition to the multiple impaction, the change of solids concentration in the liquid also contributes to increasing the collecting efficiency.

The pressure drop for impingement baffle scrubbers ranges from 2 inches WG to 8 inches WG depending on the number of stages, size, and numbers of perforations and baffles. As would be expected, increased number of stages and smaller perforations result in higher pressure drop and subsequent higher efficiency. An impingement baffle scrubber is shown in Figure 5-4.

The impingement baffle scrubbers are normally used in lime kiln applications. The efficiency is, however, not high enough to comply with today's air quality requirements. For upgrading of existing impingement baffle scrubbers, it is possible to install a Venturi and an elbow ahead of the scrubber and to use the scrubber shell as a cyclonic separator. The inlet to the scrubber must, in such a case, be changed to enter the shell tangentially, and all the internal parts have to be removed. When estimating the cost of altering an impingement baffle scrubber to a Venturi, the cost figures for Venturi scrubbers may be used. The actual cost will vary largely, depending on local conditions.

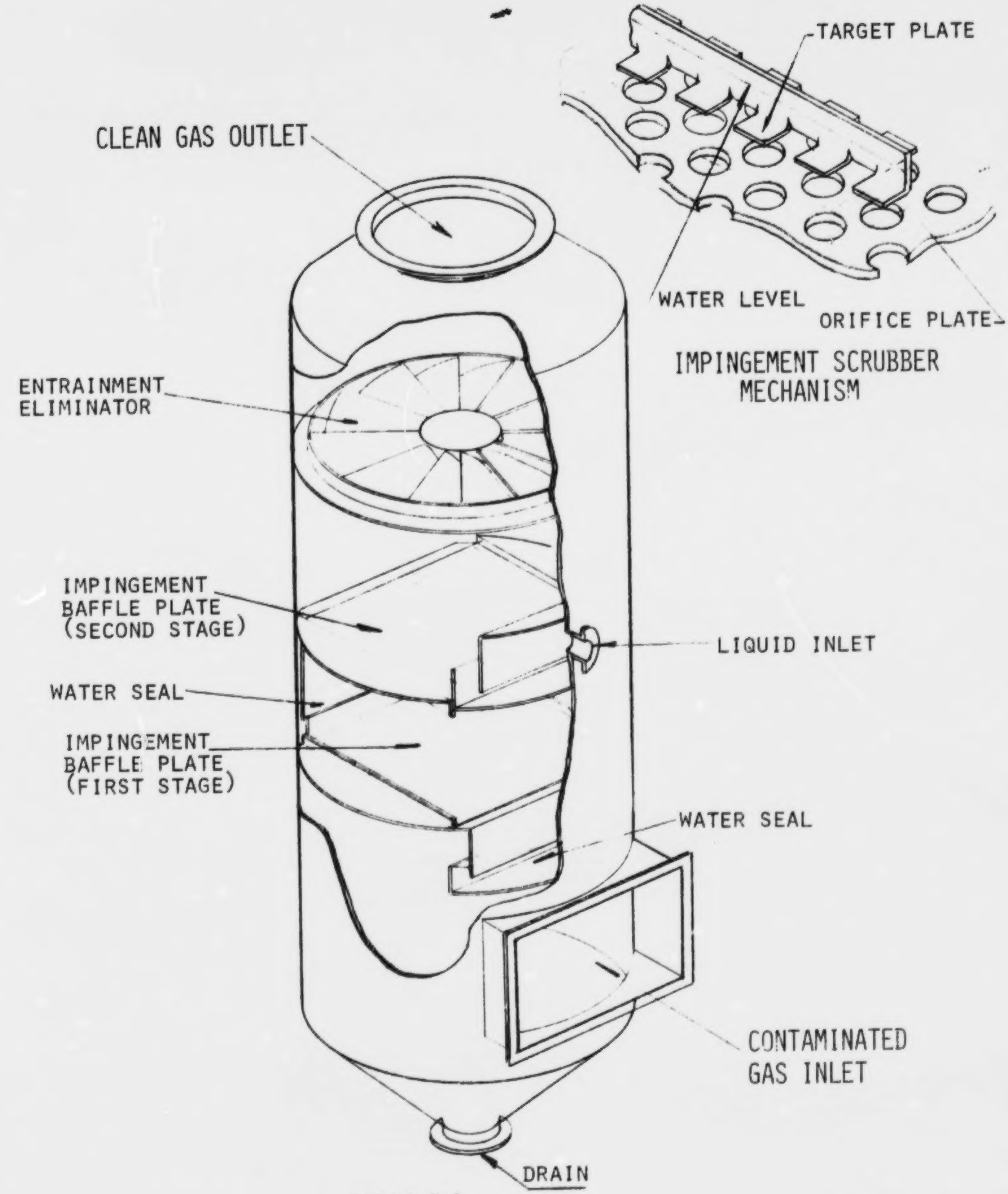

TARGET PLATE
CLEAN GAS OUTLET
WATER LEVEL
ORIFICE PLATE
IMPINGEMENT SCRUBBER MECHANISM
ENTRAINMENT ELIMINATOR
IMPINGEMENT BAFFLE PLATE (SECOND STAGE)
LIQUID INLET
WATER SEAL
IMPINGEMENT BAFFLE PLATE (FIRST STAGE)
WATER SEAL
CONTAMINATED GAS INLET
DRAIN

FIGURE 5-4

IMPINGEMENT SCRUBBER

5.2.5 PACKED TOWER SCRUBBERS

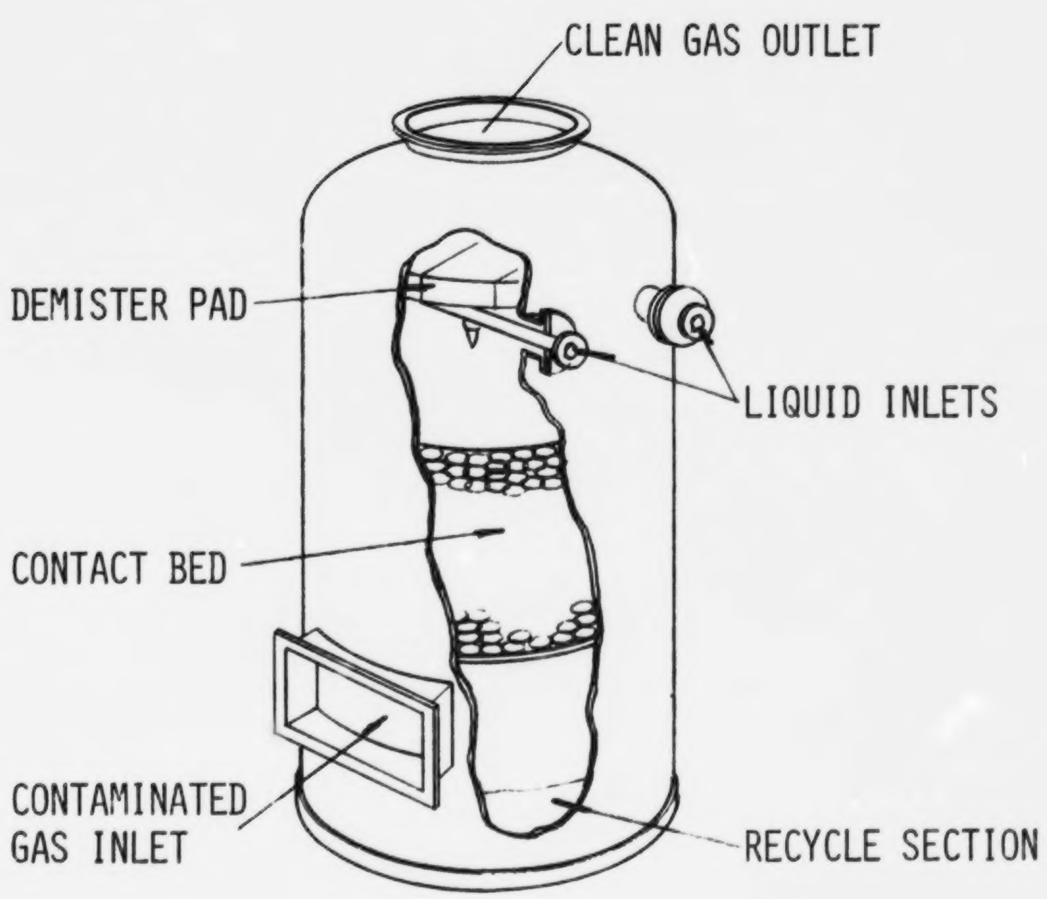
The packed tower scrubber is used primarily as a gas absorber. Its use for collection of solid particulates is limited because it is rather inefficient for particles under five microns and is subject to becoming plugged because of dust build-up. It is, however, an excellent device for absorption of such gases as HCl, SO_2, Cl_2, H_2S, and NH_3. Other advantages of the scrubber are simple design and low manufacturing cost.

The scrubber consists of a vertical cylindrical shell with the gas inlet at the bottom and the outlet at the top. Above the gas entrance is a packed section consisting of four or more feet of packing material. This packing may consist of Raschig rings, Pall rings, saddles, et cetera, made from stoneware, ceramic, or polypropylene. Water is distributed uniformly over the packing by means of low pressure spray nozzles or weirs located above the packing. Normally, there is a mist eliminator above the spray nozzles to prevent entrainment of liquid in the clean gas leaving the tower. Normally a reservoir is located at the bottom of the scrubber for direct recirculation. A typical packed tower scrubber and packing are shown in Figure 5-5.

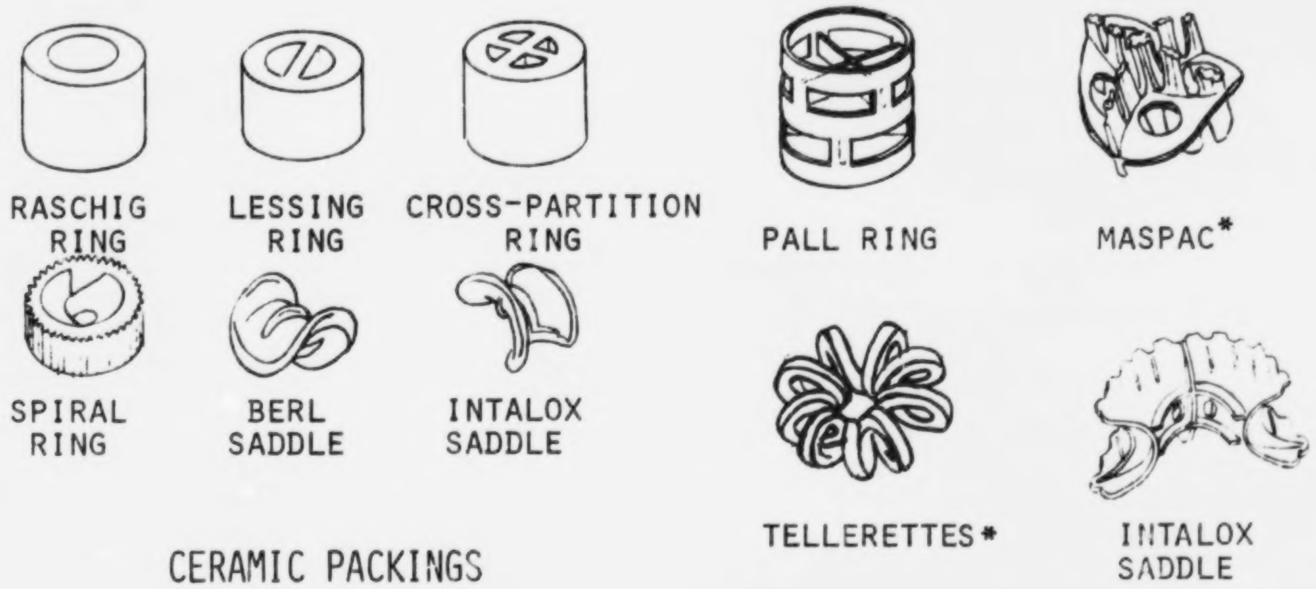
The contaminated gas enters at the bottom of the tower and moves upward through the packing counter-current to the scrubbing liquid. The packing forces the gas to follow a tortuous path over the contact surfaces and interstices creating intimate contact with the descending liquid.

The pressure drop over the tower depends on the height of packing but is normally in the order of 1 - 2 inches WG.

As mentioned before, the packed tower is an excellent gas absorber, but less suitable for dust removal due to dust build-up. When dust is present in the gas, special attention has to be given to the selection of packing material. A larger size packing; i.e., 4-inch Pall rings instead of 2-inch should be chosen in order to prevent plugging. The contact surface will thereby be reduced and the gas absorption efficiency will drop. This can be compensated for by increasing the height of the packing but the plugging tendencies will increase simultaneously.


CLEAN GAS OUTLET
DEMISTER PAD
LIQUID INLETS
CONTACT BED
CONTAMINATED GAS INLET
RECYCLE SECTION

COUNTER-FLOW PACKED TOWER SCRUBBER


RASCHIG RING
LESSING RING
CROSS-PARTITION RING
PALL RING
MASPAC*
SPIRAL RING
BERL SADDLE
INTALOX SADDLE
TELLERETTES*
INTALOX SADDLE

CERAMIC PACKINGS

PLASTIC PACKINGS

FIGURE 5-5

5.2.6 MECHANICAL COLLECTORS

There are a great number of different mechanical collectors in use throughout industry. The most common ones within the wood pulping industry are the large diameter cyclones and the multi-tube collectors. This description will be limited to these two types.

Cyclone dust collectors are of cylindrical or conical type, utilizing centrifugal and gravitational forces for separation of dust particles from a gas stream. The dust laden gas enters the collector tangentially either directly or via an expanded involute section where the dust particles are subjected to the separating forces. The centrifugal force drives the dust particles to the collector wall; the gravitation drives the concentrated dust downward to the cone outlet; and the dust is discharged into a collection hopper while the cleaned gas flows upward in an inner vortex to the gas outlet tube.

Two basic types of cyclone collectors are available--the tangential inlet type shown in Figures 5-6 and 5-7 and the axial vane type. The former is sometimes referred to as large diameter cyclone, or cyclone collector, and the latter one as a tubular collector. In the tangential inlet type, the gas enters the cyclone through a straight tangential, helical, or involute inlet section. Axial vane units employ inlet vanes to provide the spiraling motion to the dust laden gas stream.

Many sizes and designs of cyclone collectors can be provided to meet specific dust collection problems. The units may be installed in single or multiple arrangements, in parallel or in series. Cyclone collectors are generally suitable for separating solid particles in size ranging from about 3 microns to 200 microns. They can, of course, be used for larger particle sizes also.

For a given cyclone collector, the overall collecting efficiency can be determined if certain parameters are known. These parameters include dust particle size distribution, density, concentration, and other properties of the dust. Gas temperature, pressure, moisture content and gas composition, as well as pressure drop limitations and local

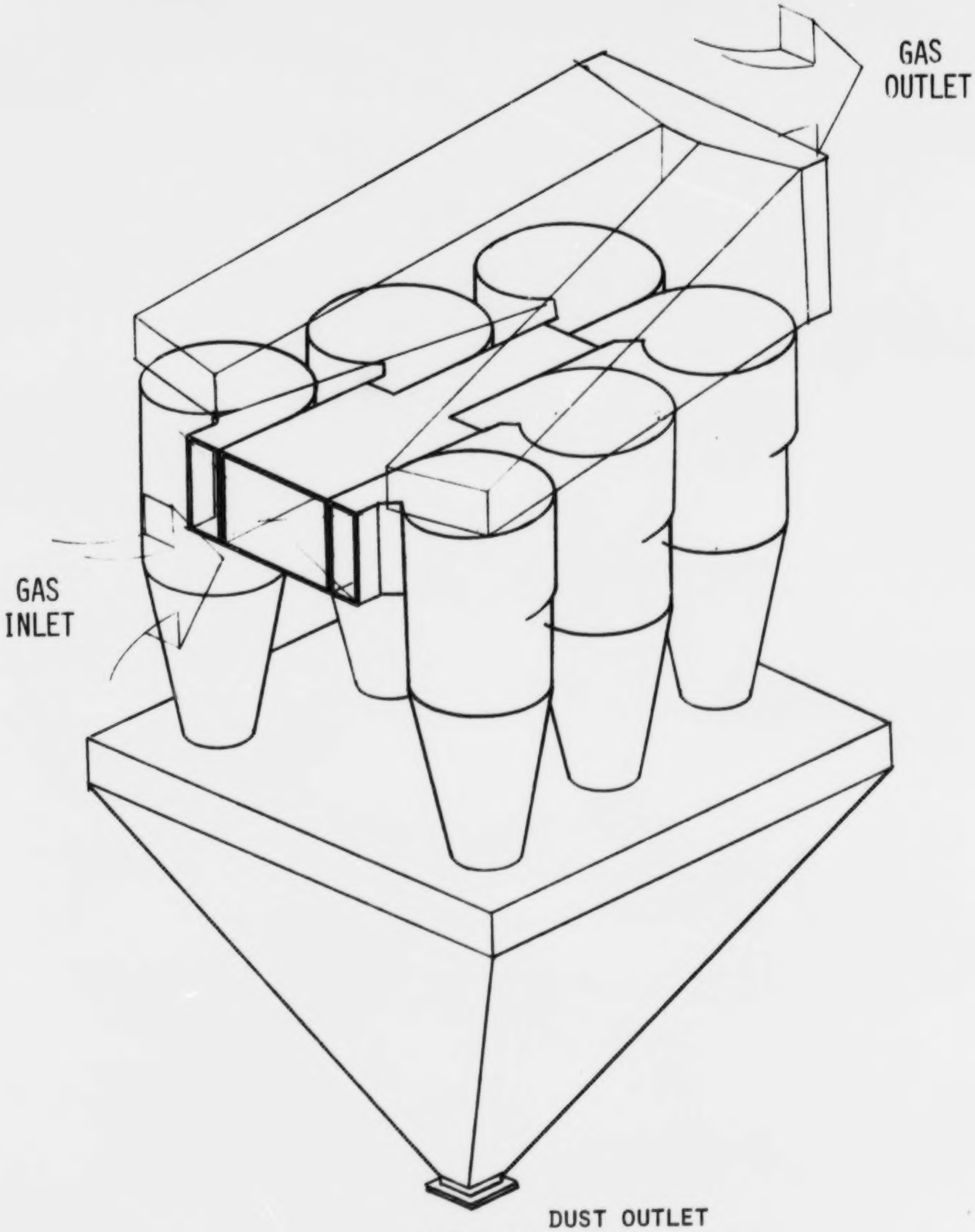
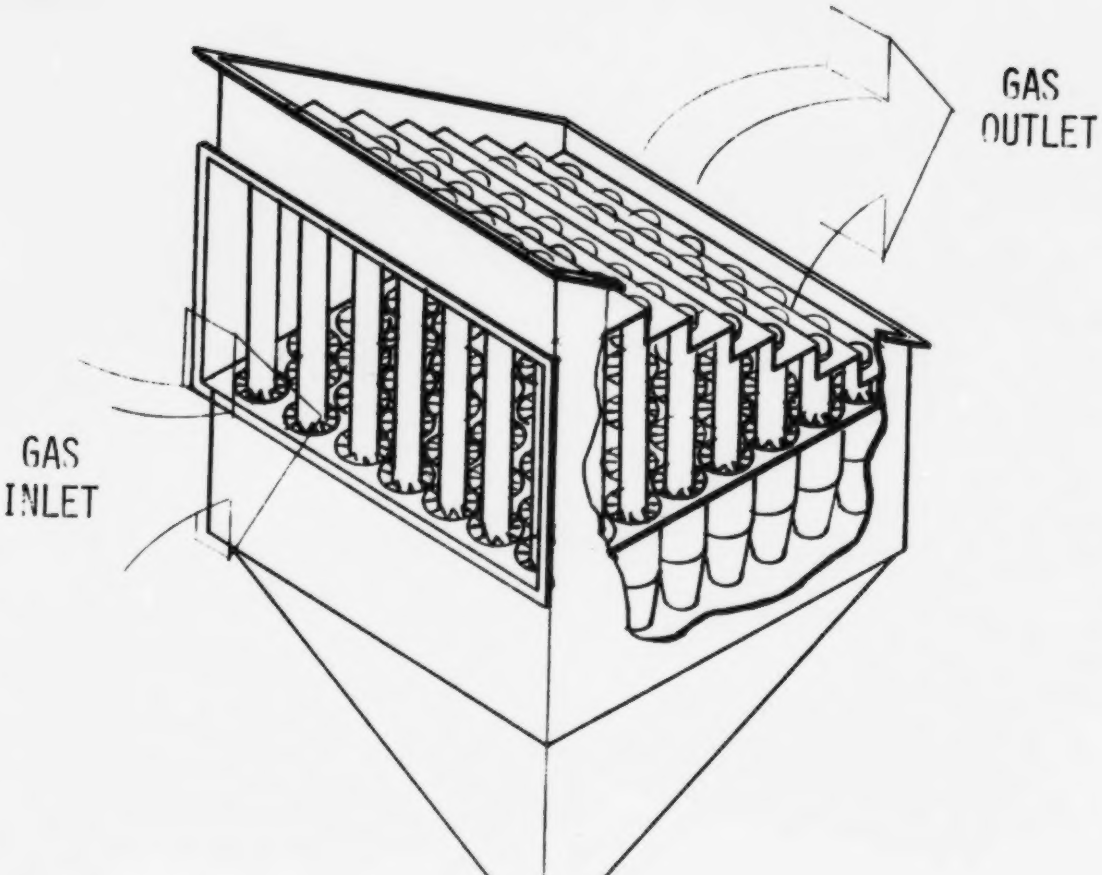
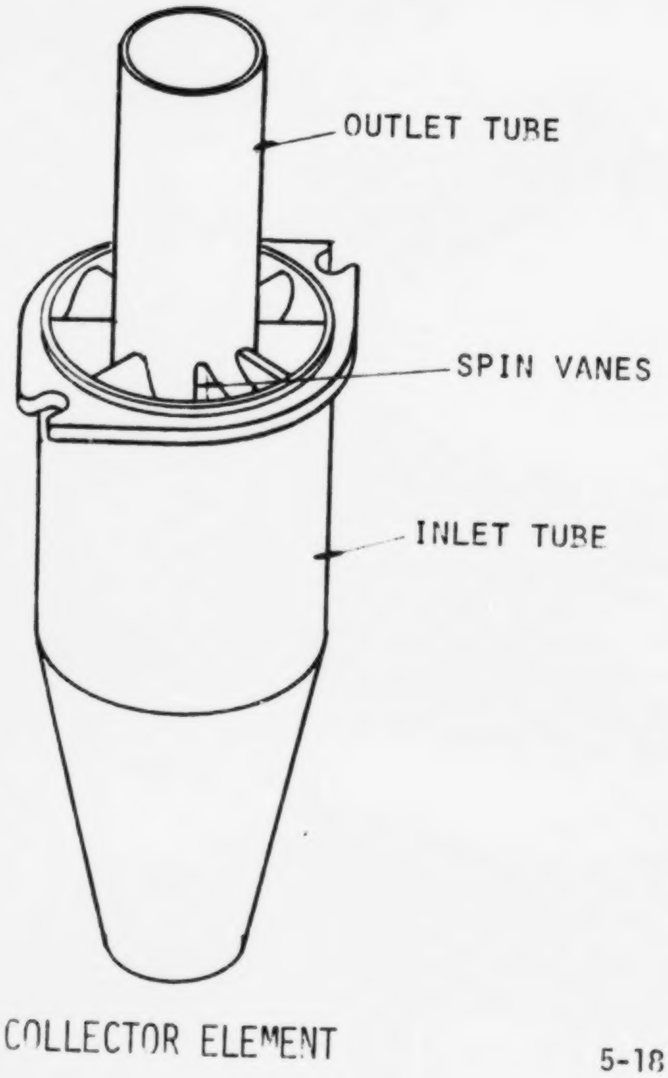

GAS OUTLET
GAS INLET
DUST OUTLET

FIGURE 5-6

LARGE DIAMETER CYCLONE COLLECTOR
STANDARD ARRANGEMENT


GAS OUTLET
GAS INLET

MULTI-TUBE COLLECTOR


OUTLET TUBE
SPIN VANES
INLET TUBE

COLLECTOR ELEMENT

FIGURE 5-7

conditions, are also factors of consideration. The efficiency increases with increased dust particle size, density, and concentration. Collection efficiency increases with increased pressure drop across the cyclone. The collector can be designed for pressure drops of less than 2 inches WG for large diameter cyclones and for 2 - 7 inches WG for the smaller diameter (high efficiency) units.

Increasing temperature will decrease the efficiency of the cyclone. However, up to 700°F, the influence of the gas temperature is not great. For the same pressure drop, increased gas viscosity or density will lower the collecting efficiency.

As particle size, dust density, and inlet gas velocity decreases the efficiency tapers off. Overall efficiency of the collection system can be increased by arranging the cyclones in series. By reducing the diameter of the cyclone the centrifugal forces increase, thus increasing the efficiency. The gas volume capacity will, however, simultaneously be reduced and the number of cyclones has to be increased. This principle is utilized in the multi-tube collectors where the tube diameter is from 12 inches or less up to 24 inches. Batteries of tubes are mounted in the same casing and high efficiencies are attainable.

Since high efficiencies require high radial gas velocities, abrasion is often a problem for cyclone collectors. Proper selection of material of construction is imperative. Materials in use include carbon steel, low alloy steels, aluminum and special materials. Abrasion resistant linings in castable form or brick linings can also be used.

5.2.7 BLACK LIQUOR OXIDATION

Black liquor oxidation is practiced for the purpose of oxidizing the sodium sulfide in the liquor. This is accomplished by reacting the sodium sulfide with oxygen in the air to form sodium thiosulfate. This compound is relatively stable and will not break down in passing through the direct contact evaporator. Sodium sulfide reacting with the carbon dioxide and sulfur dioxide in the flue gases is the cause of hydrogen sulfide emissions. The chemical reactions of sodium sulfide (unoxidized liquor) and CO_2 and SO_2 in the flue gases are as follows:

$$Na_2S + CO_2 + H_2O \rightarrow Na_2CO_3 + H_2S$$

$$Na_2S + SO_2 + H_2O \rightarrow Na_2SO_3 + H_2S$$

Black liquor oxidation can be applied either to weak black liquor or to concentrated black liquor. The choice is dependent to a considerable extent on the characteristics of the liquor. Usually weak liquor oxidation is used on non-foaming liquor and concentrated liquor oxidation on foaming type liquor. Regardless of the type of oxidation, it is essential that complete oxidation of the sodium sulfide takes place, if hydrogen sulfide emissions from the direct contact evaporator are to be prevented.

Lately, investigations have been conducted into sodium sulfide reversion taking place after weak liquor oxidation. Instances are known to exist where, although there is 99 plus percent oxidation during weak liquor oxidation, the concentrated liquor shows the presence of sodium sulfide. One solution is to add a second system of concentrated liquor oxidation. Among those who have installed weak liquor oxidation and who are having sodium sulfide reversion, there are suggestions that concentrated liquor oxidation might be the correct installation (14).

There are three types of oxidation systems:

1. Packed Towers
2. Bubble Tray Towers
3. Air Sparged Reactors.

Packed Towers

Packed towers have been applied primarily to weak black liquor oxidation. The principal advantage of a packed tower lies in reduced power costs. The pressure drop through a packed unit is usually about 1 to 2 inches WG, and a tower for a 500 ton-per-day mill will require about 20 hp to operate the required air blowers. In regions of high power costs, this is an important factor in selecting an oxidation unit. The disadvantage of this unit is that it has plugging tendencies and has lower oxidation efficiency.

Bubble Tray Towers

The bubble tray oxidation tower is used for weak black liquor oxidation only. The liquor is pumped out on a perforated steel plate, under which air is blown from a fan. The air passes through the perforations and bubbles through the liquor. The liquor height on the plate is normally four to six inches and the liquor makes several passes over the plate. In order to accommodate larger flows of liquor, a number of these aeration chambers (bubble trays) are connected in parallel and stacked on top of each other.

The liquor, air, and foam are discharged into a foam tank, where mechanical foam breakers convert the foam to liquor. Certain liquors with low foam characteristics may not require foam breakers if the tank is large enough. The air leaving the system through foam breakers carries some entrained liquor. This liquor is separated from the air in a cyclone and returned to the foam tank, while the air is exhausted to the atmosphere. This air stream, although relatively small, contains some hydrogen sulfide, methyl mercaptan, dimethyl sulfide, and dimethyl disulfide.

The system operates at 25 - 35 inches WG pressure drop. The foam breakers draw approximately 10 BHP each. Including necessary pump capacity, the total power consumption runs about 0.25 BHP/GPM. The air to liquor ratio is 15 - 20 CFM/GPM and a retention time of 4 - 6 minutes is required on the perforated plates.

Air Sparger Reactors

The air sparged reactors are of two types--air sparging with agitation and without agitation.

Air Sparger With Agitation. Air sparging with agitation is used for both weak and concentrated liquor. This system requires more electric power, it is more complicated, and the equipment cost is higher than for the non-agitation system. The liquor is pumped into a tank where an air header comes in centrally at the top. The header conveys the air into the sparger located approximately six feet from the bottom of the tank. The sparger has a number of arms extending radially from the header and each arm has a number of branches with aeration nozzles. The sparger is submerged ten to fifteen feet in the liquor depending on the desired retention time. The incoming liquor is distributed above the surface in a system of pipes and nozzles. In addition to being evenly distributed, the liquor helps to beat down the foam floating on the surface.

Air Sparger Without Agitation. Air sparging without agitation is used exclusively for concentrated black liquor oxidation. This system operates at about 10 psig air pressure and draws approximately 1.5 - 2.0 BHP/GPM of liquor. The air to liquor ratio is 20 - 30 CFM/GPM and the retention time 2 - 3 hours. Systems with agitation use more horsepower, but should otherwise be comparable.

Effect of Batch and Continuous Digesters

Mills with continuous digesters usually experience higher sodium sulfide loadings than mills with batch digesters. This is attributed to the fact that batch digester systems expose the liquor to more contact with air (primarily during washing) than continuous digester systems, thus resulting in more oxidation of the liquor. For batch digester systems the sodium sulfide content in the weak black liquor storage tanks may be 50 percent of that in the white liquor storage tanks. This condition must be considered when selecting oxidation systems.

Summary

As stated previously, foaming is a major consideration in the selection of an oxidation system. Foaming depends on the wood species, which will contain varying amounts of resin and fatty acid salts. These compounds will create a foam, that in some cases will preclude the oxidation of weak liquor. The foaming is less pronounced in strong liquor. Especially severe is the foaming of liquors from southern pine. The addition of fuel oil or kerosene to the liquor to reduce the foam helps in some cases. Foaming characteristics of the liquor influences the geographical locations of the various oxidation systems. In general, the air sparger type is located in the South due to the resinous content of the southern liquors. Weak liquor oxidation has found extensive application in western United States. The mid-west and northeastern United States have installed primarily weak liquor oxidation systems; however, (1) only one installation is reportedly operating at a high oxidation efficiency. Currently, the general trend in oxidation systems appears to be directed towards more acceptance of the air sparger type for concentrated liquor oxidation.

5.2.8 ORIFICE SCRUBBER

The orifice scrubber is a collection device consisting of a restricted air passage partially filled with water. The resulting dispersion of the water causes wetting of the particles and their collection. Pressure drop is comparable to cyclonic scrubbers.

5.2.9 MESH PADS

Mesh pads are collection devices composed of material such as knitted wire mesh. Dust and liquid droplets are collected on the pads.

A spray washing system is provided for back washing the mesh pads to remove accumulated solid particulate. In normal operation, a maximum pressure drop of approximately 0.2 inches WG is maintained by periodic operation of the spray system.

5.2.10 TALL STACKS

Tall stacks have been recognized by many authorities as a method which may be used to reduce the concentration of pollutants at ground level. Of course, the use of the tall stack will not in itself reduce the amount of pollutant being discharged into the atmosphere (17).

In discussing tall chimneys, Carlton-Jones and Schneider (13) state that "A chimney's air pollution control function is to discharge gases at a point high enough so that the maximum concentration of odors and toxic components at ground level is within acceptable limits. It is seldom practical to rely on chimneys alone for solving particulate pollution problems. For most odors and gases, however, chimneys may be a practical answer."

While there is no such thing as a standard chimney, the cost curves in Figure 5-8 are a rough estimate of chimney cost. These costs are generally based upon the following considerations:

Concrete Outer Shell
Free Standing Inner Liner Constructed of Acid Resistant Brick and Mortar
Inner Liner for the Entire Chimney Height
Foundation Allowances Included
Platforms and Ladders
Lightning Protection System
Obstruction Lighting
Wind Velocity of 100 MPH
Indirect Cost of 30 Percent

These chimney costs are based upon data from Carlton-Jones and Schneider (13) and have been updated to reflect January 1969 costs. Of course, these rough estimates will be influenced by many factors including geographic area, soil conditions, design wind velocity, labor costs, and existing physical limitations at a specific mill.

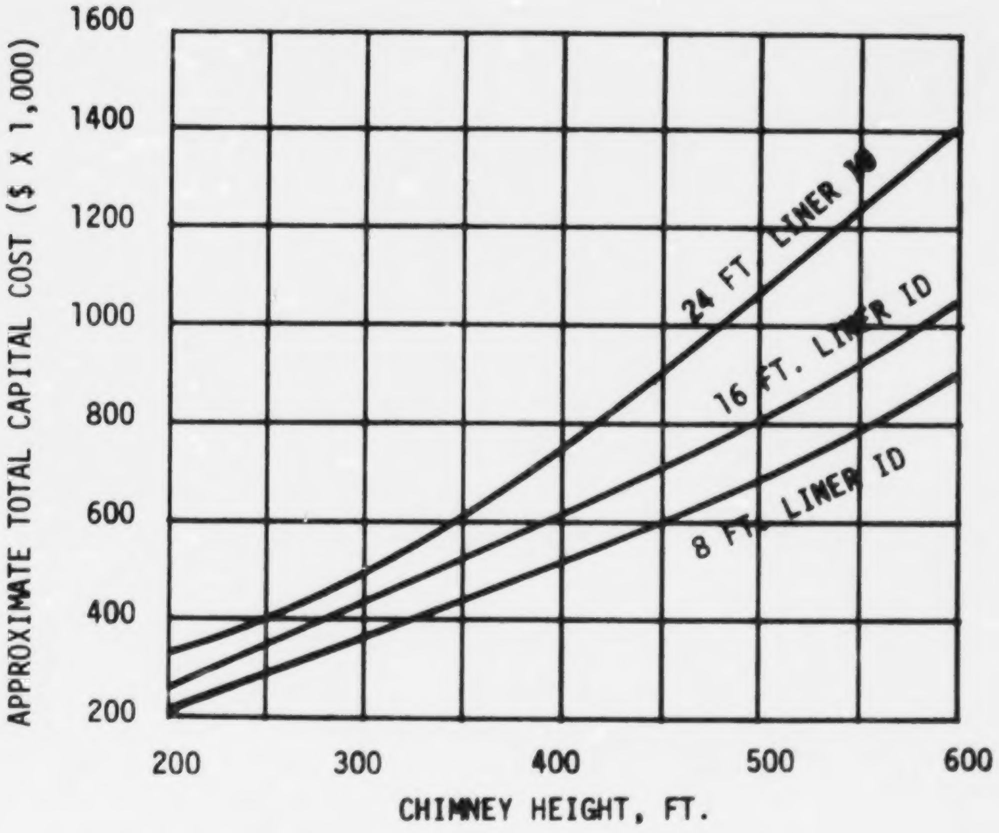


FIG. 5-8 CHIMNEY TOTAL CAPITAL COSTS

5.3 APPLICATION, COST, AND EFFECTIVENESS OF PRESENT CONTROL METHODS

5.3.1 PURPOSE OF THIS SECTION

A number of control techniques for gases and particulates has been developed and applied in the wood pulping industry in recent years. In order to provide a basis for comparing the cost-effectiveness of alternative solutions to controlling emissions from a single source and for evaluating the most effective complete system, engineering estimates are essential.

5.3.2 BASIS OF SELECTION OF METHODS

This chapter of the report is concerned with control methods presently in use in the wood pulping industry. Therefore, only control methods which have been applied successfully in at least one United States mill for at least one year are included in this chapter. The only exception to this selection basis is the inclusion of capital costs for new recovery system designs which have a great potential for reducing odorous emissions and were included because of the interest expressed by the industry.

Because these methods represent United States applied practice, techniques are not described for some sources and for some emissions. The reader is referred to Chapter 6 for a review of promising potential control methods.

5.3.3 DEFINITION OF EFFECTIVENESS

A concise definition of effectiveness of an emission control method is difficult to obtain because of the many factors to be considered. In addition to the physical efficiency of a device, its reliability, application, operating conditions, degree of maintenance, and energy reliability must be considered. For the purpose of comparing control methods, this report discusses the effectiveness of the methods in terms of the following categories:

1. Particulate removal
2. Sulfur oxide removal
3. Total reduced sulfur (TRS) removal
4. Operation

Particulate Removal. The particulate removal of control methods has been evaluated in terms of "Mean Annual Operating Efficiency"

(A.O.E.) which is a function of a number of variables including the "Guaranteed Efficiency." The "Guaranteed Efficiency" is defined as the manufacturer's guaranteed efficiency which is calculated in accordance with Industrial Gas Cleaning Institute (IGCI) procedures. IGCI specifies a test procedure (15, 16) for determining the collection efficiency of particulate collectors, similar to ASME Power Test Code 21, which is used by most equipment manufacturers for guaranteeing collector efficiencies. The method is based on a determination of the average particulate concentration at the inlet and outlet of the collector. Isokinetic sampling at specified locations in the ducts for specified times is required at steady operating conditions. The particle collector is specified as a filter of any material and form which has a collection efficiency in excess of 99.0 percent for particulates of the approximate size distribution to be encountered during the test. Collectors other than filters are acceptable if it can be demonstrated that they have an efficiency exceeding 99.0 percent at test conditions.

The "Mean Annual Operating Efficiency" (A.O.E.) is defined as:

$$\frac{\text{Total Annual Particulate to Control Method - Total Annual Particulate Leaving Control Method}}{\text{Total Annual Particulate to Control Method}} \times 100$$

The A.O.E. is a value that is difficult to obtain in actual practice. The contractor has used past experience, operating data which has been obtained from a number of mills, and theoretical calculations to derive A.O.E.'s which are theoretical, but are considered typical of the particular control method.

Sulfur Oxide and Total Reduced Sulfur (TRS) Removal. The control methods are also evaluated in terms of their effectiveness in removing sulfur oxides and TRS. Since data are limited in this area, the discussion is generally limited to whether the control methods have any potential for removing the sulfur oxides and TRS.

Operation. The effectiveness evaluation also includes a comparison of the operation of control methods. Included in this comparison are such things as ease of maintenance, corrosion potential, and plume development.

For each source, the most effective method is chosen based on the effectiveness evaluations and the control method cost.

5.3.4 BASIS FOR ENGINEERING COST CALCULATIONS

Capital costs and annual costs have been calculated, based on assumptions and values that may be considered reasonably typical for the industry in general. All of these costs can be expected to vary, depending on various factors such as the particular company's accounting policies and financial situation at that time, market conditions, space limitations, geographical location, and particular mill operating considerations. All costs are based on January 1, 1969, prices.

Utility costs and chemical costs which have been used are tabulated on page 5-32.

Costs for control methods are calculated for significant sources of emissions. Where older equipment exists, control methods are considered to reduce that particular source to the minimum emission based on current technology. If a further reduction in emissions may be necessary, it is possible that new equipment designs would have to be installed to replace the existing equipment. However, the consideration of replacing existing process equipment with new equipment is beyond the scope of this study; with the exception of New Recovery Systems included in Section 5.3.6.1.6.

Capital Cost

Capital cost has been calculated for each control method based on a typical arrangement as shown in this section. The capital cost has been calculated on an arrangement designed to minimize installation downtime. A description of the "Capital Cost" is included on page 5-29.

"Loss of Production" cost has not been included since this will vary from mill to mill. However, "Loss of Production" cost may be very significant and must be included for individual mill applications. For example, a two-week shutdown at a 500-ton mill might result in the following loss to the mill:

Loss of Profit:	
$20/Ton (profit after taxes) x 500 T/D x 14 days =	$140,000
Loss of income to pay for continuing expense of general overhead and administration--approximately the same as Loss of Profit =	140,000
Approximate Loss to the Mill	$280,000

The above calculation does not include production personnel wages, which may also be a "Loss of Production" cost.

Annual Cost

Annual cost has also been calculated in accordance with the description on pages 5-30 and 5-31.

As described under "Depreciation and Interest (Capital Recovery)," page 5-31, an interest rate of 10 percent has been used. After the total annual cost is calculated, applicable credits are deducted, resulting in a net annual cost or savings. This net annual cost or savings is then plotted on a curve along with the capital cost. In calculating the net annual cost, income tax credits, and investment tax credits have not been included.

Control Methods

Control methods to be considered for evaluation are discussed under each source. Where preliminary engineering evaluations and cost calculations indicate that a particular control method is not applicable, this particular control method has not been investigated further. However, the results of these evaluations and cost calculations will be reported for the particular control method.

Sketches and Flow Schematics

Where appropriate, sketches and flow schematics are included for a control method to better describe the basis for the calculations of "Capital Cost" and "Net Annual Cost."

Cost Curves

Cost curves are included for control methods which have been evaluated as being reasonably effective in reducing air emissions for an individual source. The cost curves--"Capital Cost" and "Net Annual Cost"--have been plotted versus a function of size, tonnage, or CFM.

Detailed cost back-up sheets were prepared. However, due to the voluminous quantity of this data, these detailed sheets are not included with this report.

DESCRIPTION OF CAPITAL COST ITEMS

Purchased Equipment

Cost of the emission control device and all accessories and auxiliaries required for its full operation.

Equipment Erection

The erection of purchased equipment only. The erection cost of building and foundations, piping and wiring is included elsewhere.

Equipment Foundations and Building

The installed cost of all equipment foundations, building and building foundations, floors, roof, stairs, and walkways, that are required for support, access, and enclosure of the control device.

Process and Instrument Piping

Cost of all pipe and pipe supports, erected, and including insulation and protective coating where required.

Power Wiring and Lighting

The installed cost of all power wiring and lighting is included. Wherever required, any substation, transformers, or switchgear costs are included under "Purchased Equipment."

Indirect Capital Cost

The summation of the above capital cost composes the total direct capital cost to which is added indirect cost as a percentage of direct cost. The following is a breakdown of this indirect cost:

15%	Contingency--Unforeseen conditions and items not practical to estimate
7%	Engineering--Preparation of specifications and working drawings, selection, and evaluation of equipment
1%	General Construction Overhead--Includes temporary facilities, contractual supervision, timekeeping, et cetera
3%	Start-up Cost--Loss of production not included
2%	Spare Parts
2%	Sales Tax
30%	Total

DESCRIPTION OF ANNUAL COST ITEMS

General

Any estimated annual cost may vary considerably from mill to mill. The following values may be considered typical and should provide reasonable comparisons of emission control device annual cost:

Direct Operating Cost

Direct operating cost consists of charges for:

1. Operating Labor, including Overhead. A national average hourly rate of $4.25 has been assumed. To this rate, an additional 29 percent is added for vacations, sick pay, holidays, payroll taxes, insurance, and fringe benefits. (Total $5.50 per hour)
2. Power, Electric, and/or Steam. An electric cost of $0.01 per KWH, and a steam cost of $0.65 per 1000 lbs. of steam has been used.
3. Water. A water cost of $0.13 per 1000 gallons has been used.
4. Maintenance, including maintenance labor, replacement parts, and maintenance materials.

Taxes and Insurance

Local property taxes vary widely over the country. Generally, this is applied as a millage on the assessed valuation. Assuming the property is assessed at 50 percent and 30 mills tax is applied, then the property tax would be one and one-half percent of the construction cost.

	1.5% of Capital Cost
An average insurance cost has been used for the entire pulp and paper mill.	0.5% of Capital Cost
Total Taxes and Insurance	2.0% of Capital Cost

Administrative Costs

This is an average cost applied to the overall plant and includes all salaried personnel (officers and supervisors), fringe benefits for salaried personnel, legal and other professional services, public relations, contributions, and office supplies and expenses. This does not include any marketing costs.

5.0% of Capital Cost

Depreciation and Interest (Capital Recovery)

Depreciation plus interest charges were calculated from the following formula which is one of several commonly used for this purpose. An interest rate of 10 percent was assumed.*

$$\text{Equal Annual Payment} = \frac{i(1 + i)^n \times 100\%}{(1 + i)^n - 1}$$

i = interest rate

n = life in years

SUMMARY OF CAPITAL CHARGE ITEMS
(As % of Capital Cost)

	Estimated Life of Equipment in Years		
Item	8	10	16
Taxes & Insurance	2%	2%	2%
Administration	5%	5%	5%
Depreciation & Interest (Capital Recovery)	18.7%	16.3%	12.8%
TOTAL	25.7%	23.3%	19.8%

*Much of the industry uses a rate of nearly 20 percent before taxes which reflects the fact that an investment in air pollution control equipment is a lost opportunity for profit compared with the alternative course of investing the funds in productive equipment which would produce income equal to the average current rate of return or unamortized investment for the industry (18).

UTILITY COST

Item	Fixed Charges	Operating Costs: Fuel, Labor, Maint. Etc.	Total (Rounded to Nearest Cent)
Steam $/1000 lbs. Based on 1000 BTU/lb.	$0.116	$0.535	$0.650
Electricity $/KWH	0.0023	0.0077	0.010
Water $/1000 gal.	0.104	0.046	0.150
Kraft Waste Treatment $/1000 gal. (Primary and Secondary Treatment)	0.13	0.03	0.160

These are values that may be considered typical in a general sense. All of these values vary considerably from mill to mill--any proper evaluation of control costs should be specifically calculated for the individual mill. However, the above costs should provide reasonable comparisons of emission control method for this study.

CHEMICAL COSTS
Per Ton Except as Noted

Salt Cake	$34.00 East $24.50 West	Average: $30.00
Lime (CaO)	$15.00	Average: $15.00
Sulfur	$39.00 to $42.00 per Long Ton	Average: $40.00
Soda Ash	$31.00 to $32.00	Average: $31.00
Caustic Soda	$57.00 per ton of NaOH in a 50% Solution	
Magnesium Hydroxide	$37.68 per ton 100% Solids-- Tank Car Lots, FOB, Michigan	
Chlorine (Papermakers)	$ 3.35 per 100 lbs.	

5.3.5 LIMITATIONS TO APPLYING COST CALCULATIONS

The design, arrangement, and operation of pulp mills demonstrate as many variations as there are mills. The application of control methods to mills is equally as varied as the mills themselves. In order to determine a source "arrangement" which might be considered typical, a survey of all configurations at all mills would have to be completed. Rather than attempt this magnitude of survey and data compilation, the experience and judgment of the contractor was employed to assume arrangements which might be considered reasonably typical. These arrangements are the basis for computing cost calculations for control methods applied to the various sources.

The arrangements and cost calculations are considered to be reasonably typical of pulp mills in the United States. Application of these costs to an area smaller than the United States is not intended. The costs are considered to be accurate within ± 20 percent over the entire United States. However, if these costs were to be used for one mill only, the variation might be from 70 to 200 percent of the U. S. average. For individual applications, specific and detailed costs must be calculated. For instance, the application of a stainless steel precipitator in a relatively inaccessible location would more than double the precipitator cost which has been used to calculate the average U. S. cost involving a precipitator. Therefore, these cost calculations must be used judiciously.

5.3.6 KRAFT SOURCES

5.3.6.1 Recovery Systems

The design and operation of recovery systems and auxiliary equipment has been described in detail in numerous publications (2, 3, 4, 5). This discussion is primarily limited to factors influencing air emissions from recovery systems.

Design Considerations

Kraft pulping operations yield approximately 2,000 to 3,500 pounds of dry black liquor solids per ton of pulp produced. Heating values range from approximately 6,000 to 7,200 BTU's per pound of dry solids. Three thousand pounds of solids at a heating value of 6,600 BTU's per pound is generally regarded as an average. The actual size of a recovery unit is based on the heat input (solids flow times heating value). Recovery unit rating is defined as the manufacturer's guaranteed heat input to the recovery furnace. Therefore, percent overload or percent of rated capacity is normally a percentage of the manufacturer's guaranteed heat input. For ease of reference and discussion, recovery unit operators often use the manufacturer's guaranteed solids flow to the recovery system as the rated capacity. This type of reference may be satisfactory for general discussions, but the guaranteed heat input to the unit is the precise definition of rated capacity.

Steam pressure and steam temperature have increased over the years as required by economics relating to the steam's power balance. Present pressures vary from 300 to 1500 psi. Temperatures have also followed turbine practice with present units as high as 925°F.

Furnace

The furnace or combustion chamber is one of the most important components in the recovery system from an air emission standpoint. The furnace performs the following functions:

1. Dehydrating the black liquor and burning the organic constituents with maximum combustion efficiency.
2. Reducing that portion of the chemicals present as sodium sulfate and other sodium-sulfur-oxygen compounds to sodium sulfide.
3. Melting the inorganic chemicals for continuous removal in molten form.
4. Conditioning the products of combustion to reduce chemical carryover and flue gas temperature.

Black liquor is introduced to the furnace by one or more sprays located in the furnace walls. The fundamental difference between the furnaces of the two major United States manufacturers is the method of dehydrating the black liquor.

In one design of furnace, heavy black liquor is sprayed and deposited in a partially dehydrated form in a band on the furnace walls where it remains until dry before falling to the hearth to form the char bed. The spray consists of one or two nozzles oscillated in a "paintbrush" motion to obtain uniform distribution of liquor on the walls. Usually, the spray is located in the front wall, although larger units also have additional sprays in the other furnace walls.

The other furnace design introduces liquor through a multiplicity of sprays located in two or more walls so that the droplets dry in suspension. The sprays are oscillated in a vertical plane moving from horizontal to an adjustable downward angle. A very coarse spray is maintained so that, in falling from the nozzle to the hearth, the particles of liquor are flash dried. Most units use nozzles in at least two walls, with larger ones having sprays in all four walls.

Combustion of the black liquor char begins on the hearth of the furnace. Air for combustion is supplied by a forced draft system to the reducing and oxidizing zone of the furnace. Since a reducing atmosphere is required to convert sodium sulfate and other sodium-base sulfur compounds to sodium sulfide, only a portion of the air required for complete combustion is supplied to the char bed through the lower or primary air ports. The heat released by the combustion in this zone is sufficient to liquefy the chemicals in the char and to sustain the endothermic reduction. The liquefied chemical, or molten smelt, is continuously drained from the hearth.

Air is admitted above the primary zone to complete the combustion of the volatile gases from the char on the hearth. The final stage of combustion on the "paintbrush" spray system is supported by a series of ports, at secondary and tertiary levels, which direct the air across the furnace, whereas this is accomplished in the other furnace design by air admitted through a vertical bank of ports tangentially directed into the gases.

The control of total air in relation to the black liquor flow and the proper distribution of air is essential for efficient combustion, chemical reduction, and minimum emissions. A number of tests has been conducted and published indicating that approximately 10 to 15 percent excess air is essential to obtain complete, efficient combustion and minimum air emission. Tests (6) have been conducted indicating the importance of proper distribution of primary and secondary air to the furnace.

When the furnace is operated to provide for minimum emissions (primarily minimum H_2S) to the atmosphere, the most complete combustion conditions also exist. While not proven completely at the present time, it appears that SO_2 in the flue gas is also low when the H_2S is a minimum.

Apparently, best furnace operation from the standpoint of efficient, complete combustion and chemical reduction results in minimum emission of pollutants from the furnace.

Direct-Contact Evaporator

Black liquor from the multiple-effect evaporators at 45 - 55 percent solids is not sufficiently concentrated to permit firing directly into the furnace. To raise the concentration to a level which will ensure stable combustion, present U. S. practice requires the removal of additional water by means of a cascade evaporator, a cyclone evaporator, or a Venturi evaporator-scrubber. In practice, the liquor is concentrated to the range of 60 - 70 percent solids before firing.

The direct contact evaporator has been identified as a major source of H_2S emission in the recovery system. The amount of H_2S emission from the evaporator varies depending primarily upon the sodium sulfide concentration in the liquor, the pH of the liquor, and the degree of gas-liquor contact, and the amount of H_2S emissions coming from the recovery furnace.

Corrective action for the prevention of H_2S emission from the direct contact evaporator is either oxidation of the liquor or conversion to one of the new recovery systems which eliminate direct contact between the flue gases and black liquor (See Chapter 6, New Developments.)

The new systems (See Section 6.2.1.1) concentrate the liquor to the firing range of 60 - 70 percent by the use of additional evaporation in the multiple effect evaporators (up to approximately 55 percent), and finishing in an air cascade evaporator (from 55 percent to 65 - 70 percent), or by the use of an additional multiple evaporator effect (commonly called a concentrator) for a resultant liquor concentration of approximately 62 percent.

Electrostatic Precipitators

The electrostatic precipitator is a very important part of the recovery system. The cost of installation and operation of a precipitator is offset by the recovery of chemicals for all but the highest collection efficiencies. With present emphasis on pollution control, however, the collecting efficiencies are much higher than would be justified from a recovery economics standpoint.

The gases emanating from a recovery boiler are essentially flue gases with a small amount of sulfur compounds and organic sulfur compounds. The particulate matter consists primarily of sodium sulfate, or in case of a soda mill, of mostly sodium carbonate. Originating from the high temperature combustion zone, this material is sublimed fumes and of extremely small particle size. It is very light and has an average bulk weight of only nine pounds per cubic foot when precipitated in dry form.

The gaseous constituents, the high water vapor content, and the characteristics of the chemicals recovered in the precipitator present serious corrosion conditions unless the equipment is properly designed. The temperature of the gas leaving the cascade or cyclone evaporator is normally 300 - 350°F. At temperatures below 300°F, local cold spots may cause condensation and create severe corrosion problems.

To provide greater flexibility in operation, the precipitator is often designed with two identical units in parallel. This arrangement makes it possible to shut down one chamber for maintenance or inspection and still maintain reasonable cleaning efficiency while routing the entire gas flow through the other chamber.

There are many materials of construction used for the shell of the precipitator such as mild steel, reinforced concrete, hollow or filled tile, or combinations of these materials. As indicated above, it is of great importance that local condensation be prevented by proper insulation or correct gas temperature. The material of the shell should, therefore, be selected with this in mind. Mild steel will have a short life under other than ideal conditions. This material is, of course, attractive because of its relatively low cost for manufacturing as well as erection. It is a well-known fact skilled welders and boiler makers are more readily available than tile setters. A steel construction precipitator is also relatively light weight, which is an advantage from the standpoint of foundation, structural steel, and erection. In order to extend the life of the steel shell, some corrosion resistant coatings have been used with reportedly good success. It is, however, of utmost importance that the steel plate be covered completely. A very minute hole in the coating may be the starting point for corrosion under the coating.

If a reasonable life is to be expected from a steel shell precipitator, it must be provided with good insulation. This insulation may, for example consist of a blanket insulation outside the shell followed by an air space and another blanket insulation. The exterior is then usually covered by aluminum panels. Hoppers or wet bottoms are usually insulated with blankets without intermediate air space.

Concrete has somewhat better corrosion resistance than steel and is often used for roof sections. For outer walls of the shell where the temperature tends to be lowest, concrete is normally used only in locations where suitable tile or skilled tile setters are not available. Some years ago, great claims were made for so-called hollow tile due to its superior thermal insulating capacity. Many cases of seepage of salt cake into the hollow spaces of the tile have been reported and filled tile is now generally considered a more practical choice.

The bottom of the precipitator can be designed in many different ways to meet various operating conditions, plant requirements, and mill preferences. There are two basic designs of precipitator bottoms--dry and wet.

The dry bottom can be constructed as a flat bottom collecting chamber underneath the precipitator zone. From there the collected material is removed by means of a drag scraper and screw conveyer and finally discharged into a rotary valve. Trough hoppers with screw conveyers and rotary discharge valves are preferred by some mills. The dry bottom may be of some advantage when used in conjunction with recovery boilers where the flue gas must not get in contact with the black liquor.

The wet bottom design allows collected material to be brought back immediately and continuously into the black liquor flow to the direct contact evaporator. The liquid surface is located below the precipitator electrodes at sufficient distance to prevent arcing. Dust falling down into the liquor is kept in suspension by motor driven agitators or pump mixers. Contact between unoxidized black liquor and the flue gases may create a source of H_2S emission.

Sneak-by of gas in wet or dry bottom precipitators is minimized by means of fixed baffles extending from the precipitation zone down to the bottom.

Possible uneven distribution of the gas flow in the inlet end of the precipitator is corrected by perforated distribution plates or adjustable baffles. Sometimes the conditions call for vibrators or rapping mechanisms to keep the baffles or the distribution plates clean.

There are different designs of supports for the emitting electrode system. Some designs are not exposed to dust build-up in the support insulators. Build-up on the inside of an insulator can lead to arc-over and reduction of collecting efficiency and eventually to cracking of the insulator which in turn may require a precipitator shutdown. Support insulators fabricated from alumina have proven themselves to withstand substantial amounts of arc-over without cracking. Arcing is easily detectable from instruments in the control cabinet. At such occurrences, it is necessary to shut down the power in the precipitator long enough to allow cleaning of the insulators before the precipitator can be put into operation again. Under certain difficult conditions, it is sometimes necessary to blow hot air into the support insulators to prevent the excessive dust build-up.

The high voltage power units should be located as close as practical to the high voltage connections of the precipitator in order to shorten the length of the high voltage conductor. It is, therefore, common to locate the transformer and the rectifier unit directly on the precipitator roof. However, it is recommended that the control cabinet be located in the main control room, thus enabling the operators to keep close control of the operation of the precipitator. Another advantage is, of course, that the control cabinet is more protected in the control room than out in the mill.

An estimated 75 percent of all new installations of electrostatic precipitators are atop the recovery boiler. This preserves valuable ground space for future expansion of the boiler. The location of the precipitator on top of the boiler also reduces the required stack height and may also possibly reduce the draft loss. An important advantage is the reduction of horizontal duct runs which are subjected to dust fall-out with subsequent maintenance problems.

Fiberglass reinforced plastic is becoming a popular material of construction for ductwork and stacks because of its ability to withstand corrosion. This material is also lightweight and easy to erect.

The I.D. fan is usually located ahead of the precipitator. The reason for this is that for roof-mounted precipitators, which is the most modern installation, the duct work will be less complicated and less expensive. The fan is, however, more exposed to erosion and dust build-up in this location. In order to overcome the build-up, the fan is normally equipped with soot blowers, which automatically provide for 10 - 15 minute cleaning periods at certain intervals. By having the fan before the precipitator, the dust generated during the steam cleaning can be captured. The disadvantage of this fan location is, however, that the precipitator operates under positive pressure and leakage can easily occur around inspection doors and flange connections. On the other hand, when the precipitator operates under negative pressure, cold air can leak in and cause cold spots and local corrosion. With the high efficiencies of modern precipitators, there is normally no problem with build-up on fans located subsequent to precipitators and steam blowers are normally not necessary. The steam may, under certain conditions, cause corrosion.

Rebuilding Existing Precipitator

Sometimes the question is brought up of how much the efficiency can be upgraded of an existing precipitator, if the internals and the electric system be completely replaced in the existing shell. This approach has to be evaluated for each individual case, but is seldom feasible due to the following reasons:

The collection efficiency attainable will be slightly higher than the original guarantee efficiency, which would not be in compliance with today's air pollution codes. Addition of a new precipitator in series with the existing one would bring the overall collection efficiency to an acceptable level with higher equipment cost. The down time, which in the first case would be considerable, would be kept at a minimum with the add-on concept. The overall cost would most certainly be lower for the add-on alternative.

Scrubber Installed Subsequent to Precipitator

In order to improve the collecting efficiency of an electrostatic precipitator or to prevent "snow-outs," a scrubber is sometimes installed subsequent to the precipitator. The snow-outs sometimes occur when the gas velocity through the precipitator is too high (overload conditions). Under such conditions, the dust re-entrains into the high gas velocity. Too frequent or vigorous rapping can also cause snow-outs. Despite very high efficiencies, some dust will always escape the precipitator. This dust has a tendency to accumulate on the inside walls of the stack and will eventually be torn off in large flakes and thrown out the stack.

There are two reasons for installing a scrubber subsequent to the precipitator:

1. The combination of precipitator/scrubber may, in some cases, give a more economic recovery unit than a precipitator alone for the same overall efficiency.

2. The scrubber (normally of low energy type) will essentially eliminate the snow-out problems at an overloaded or deteriorated precipitator.

A bypass duct is sometimes installed around the precipitator so that repair and maintenance on the precipitator can be done without boiler shut-down. The scrubber will give the same or better collecting efficiency when the precipitator is bypassed. The overall efficiency will, of course, decrease.

Often corrosion is a great problem at these scrubbers. The gas will, in most cases, reach saturation and, due to the presence of sulfur dioxide, carbon steel construction is normally precluded. In some cases where the timber has been floated in salt water, the sodium chloride content in the black liquor is rather high. This has to be given special consideration when selecting materials of construction.

Other potential disadvantages with wet scrubbers are associated with the steam plume created at the stack discharge. Such wet plumes are quite visible and may result in complaints from neighbors. The wet plume may also result in a lower plume rise which may adversely affect dispersion of the stack gases.

Low Energy Scrubbers. The low energy scrubbers are normally installed directly behind the electrostatic precipitator. The cyclonic scrubber is the most common type. Very often the scrubber is furnished with a flat bottom to facilitate the installation.

The scrubbing liquid is recirculated, but the solids content is kept below 10 percent by maintaining a proper bleed-off. At higher concentrations, the spray nozzles have a tendency to plug.

Water is most commonly used for the scrubbing medium. Addition of caustic will increase the absorption of SO_2, while weak wash may increase the emission of H_2S and other sulfur compounds.

The vessel must be corrosion resistant. In addition to manufacturing it from stainless steel and fiberglass, lined carbon steel is frequently used. The lining is normally a 1 to 2 inch concrete lining type Gunite or similar material.

The stack also has to be protected against corrosion and since a concrete lining would make the stack too heavy and necessitate reinforcing of the scrubber vessel, other means for corrosion protection have to be found. Thus, the stack is often made from carbon steel with a stainless steel sheet lining of 1/16 inch thickness. Sometimes the stack is manufactured from polyester fiberglass, which has proven to give excellent life and low maintenance. For mills with high sodium chloride content in the black liquor, the whole scrubber should be made in fiberglass.

The pressure drop over the scrubber is normally 1 to 3 inches WG and very often due to the large gas volumes, two scrubbers in parallel have to be installed. The collecting efficiency of a cyclonic scrubber for this application is 80 - 90 percent on particulate matter.

The fan is normally located ahead of the scrubber making it possible to have the stack mounted directly on top of the scrubber.

The recirculation pump is normally made from corrosion resistant material.

High Energy Scrubbers. When a higher collecting efficiency is required than is possible to obtain in a cyclonic scrubber, a high energy scrubber must be installed. The scrubber is almost exclusively of the Venturi type.

The scrubbing liquid should be water for the same reasons previously cited. Addition of caustic or alkaline chemicals will, of course, increase the absorption of SO_2. The use of weak wash may increase the emission of H_2S and other sulfur compounds. Fully oxidized black liquor can be used providing the liquor has low foaming characteristics.

The solids content in the recirculation liquid is normally kept below 30 percent. At higher solids content, the scrubbing efficiency will decrease at the same pressure drop. In other words, more energy is needed to atomize the scrubbing liquid

The scrubber has to be corrosion resistant and frequently is manufactured in stainless steel.

The pressure drop over the scrubber is between 15 and 30 inches WG, depending on the desired collecting efficiency. Efficiencies in excess of 99 percent are attainable in this pressure drop range.

The fan is normally located between the scrubber and the stack. Due to the corrosive atmosphere, the fan has to be constructed of stainless steel, be provided with split housing for easy wheel removal, and have inspection doors, water sprays, and drain connections.

The recirculation pump is normally made from corrosion resistant material.

Venturi Evaporator-Scrubber. The Venturi evaporator-scrubber is a device which both collects dust and fume from the flue gas and provides evaporation of the black liquor. By nature, the Venturi creates intimate contact between the flue gas and the scrubbing medium thereby causing almost instantaneous saturation of the gas and at the same time evaporation of the scrubbing medium. If the scrubbing medium is black liquor, the direct contact evaporator and the fume collector can both be replaced by this evaporator-scrubber. There are two types of this device--the single stage and the two-stage Venturi evaporator-scrubber.

The first type has been used on approximately thirty recovery boilers over the years, while the latter type is a rather new development, and experience is rather limited (only one or two installations).

Both types offer some advantages. Space requirement and equipment cost are lower than for any other combination of evaporator and fume collector. The Venturi evaporator-scrubber provides higher thermal efficiency than a conventional evaporator. The temperature of the flue gas leaving a one-stage

Venturi is 180 - 190°F and from a two-stage, 160 - 170°F. In Scandinavian countries, a separate cooling loop is sometimes installed to reduce the steam plume and produce warm water. The flue gas leaving such cooling loops can be brought down to 100 - 110°F.

There are, however, some serious disadvantages with these systems. The power consumption is much higher than for a conventional system. The corrosion problem is often severe but can, of course, be overcome by proper selection of materials of construction. The advantage with low equipment cost will, however, be reduced if too sophisticated a material is chosen.

The single-stage system evaporates the liquor and scrubs the flue gas in one Venturi. The entire black liquor flow received from the multiple effect evaporators is recycled to attain 60 - 65 percent solids. The bleed-off from the system is fed to the recovery boiler to be burned. Due to the high concentration of solids in the liquor, the nozzles in the Venturi throat have a tendency to plug and the liquor is difficult to atomize. To overcome these problems, steam is fed with the liquor into the throat. This steam helps atomize the liquor and keep the nozzles open.

The Venturi which normally is used has a rectangular throat with maximum width of about twelve inches. At greater widths, the sprays cannot penetrate into the center of the gas stream and the gas will escape agglomeration with loss of efficiency as a result. Due to this limitation of width, these Venturi throats tend to be extremely long at large gas volumes. The inlet transition from the boiler to the Venturi is space consuming and difficult to erect.

The pressure drop required to atomize the liquor is increased at higher solids concentration. The scrubbing efficiency drops off drastically as the liquor concentration goes up. For a given pressure drop, water as scrubbing medium will give 99 percent efficiency. At the same pressure drop, 45 percent black liquor will give 94 percent and 60 percent black liquor can be brought up to 85 percent efficiency.

The pressure drop necessary to attain good evaporation is approximately 5 - 6 inches WG, but in order to get 95 percent scrubbing efficiency, a pressure drop of about 30 inches WG is necessary.

The schematic arrangement of this system is shown in Figure 5-9.

In the two-stage approach, (See Figure 5-10), some of the disadvantages with the one-stage system are overcome. This system uses a so-called SF Venturi, as described in Section 5.2. For large gas volumes, a modified version with an annular throat is used. The first stage has a low pressure drop, 5 - 6 inches WG, and is used exclusively as an evaporator. The total flow of black liquor from the multiple effect evaporators is introduced into this stage and is recycled until the liquor has attained 60 - 65 percent solids. The bleed-off is pumped to the recovery furnace to be burned. Some dust collection is achieved in this stage, but hardly more than in a conventional cascade or cyclone evaporator. In order to prevent build-up of black liquor solids, which would eventually impede the operation, the walls of the first stage separator are flushed with black liquor from the recycle pump. The elbow connecting the Venturi and the separator is of regular and not of the so-called flooded type. The buildup of solids is, of course, a fire hazard too.

When the gas enters the second stage, it is close to saturation and has a temperature of approximately 190°F. The pressure drop in this Venturi is about 30 - 35 inches WG depending on the desired collection efficiency. (approximately 90 - 95%). This pressure drop necessitates a high velocity. The elbow connecting the Venturi and the separator is of the flooded type to prevent excessive erosion. The scrubbing liquid is recycled to between 5 and 20 percent solids concentration. The bleed-off is fed into the first stage and collected dust is thereby automatically returned to the process. In this stage, little evaporation takes place, but the gas leaving the scrubber has a temperature of approximately 160 - 170°F and is completely saturated. The make-up can be black liquor, but excessive foaming normally precludes the use of liquor. Addition of caustic will, of course, increase the absorption of SO_2, while white liquor or weak wash will increase the emission of H_2S and other sulfur compounds. Unoxidized black liquor will also increase emission of sulfur compounds. Water is normally the best make-up and since the amount added

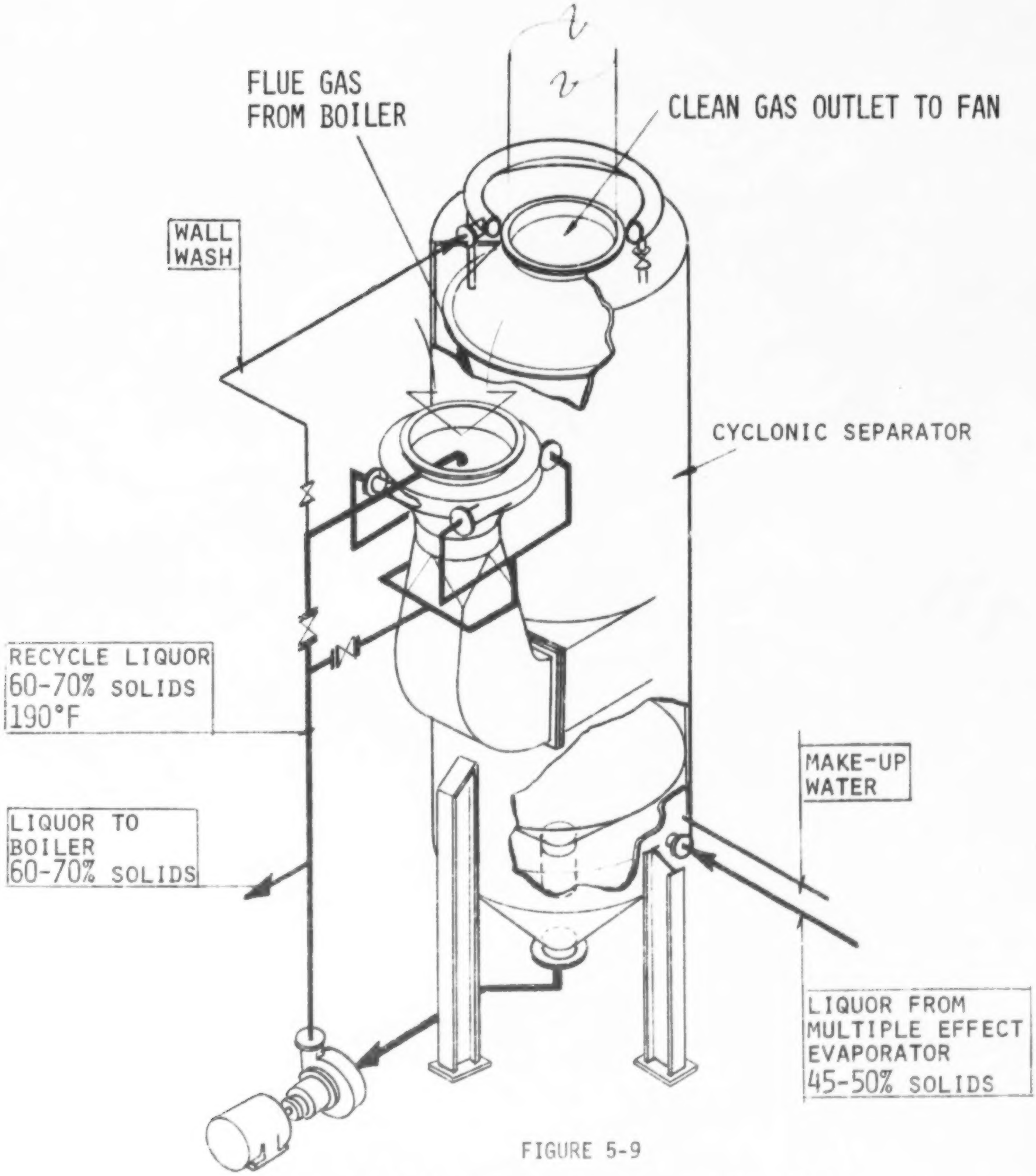

FLUE GAS
FROM BOILER
CLEAN GAS OUTLET TO FAN
WALL WASH
CYCLONIC SEPARATOR
RECYCLE LIQUOR
60-70% SOLIDS
190°F
MAKE-UP WATER
LIQUOR TO BOILER
60-70% SOLIDS
LIQUOR FROM MULTIPLE EFFECT EVAPORATOR
45-50% SOLIDS

FIGURE 5-9

TYPICAL FLOW SHEET FOR SINGLE STAGE VENTURI EVAPORATOR/SCRUBBER SYSTEM FOR RECOVERY BOILER

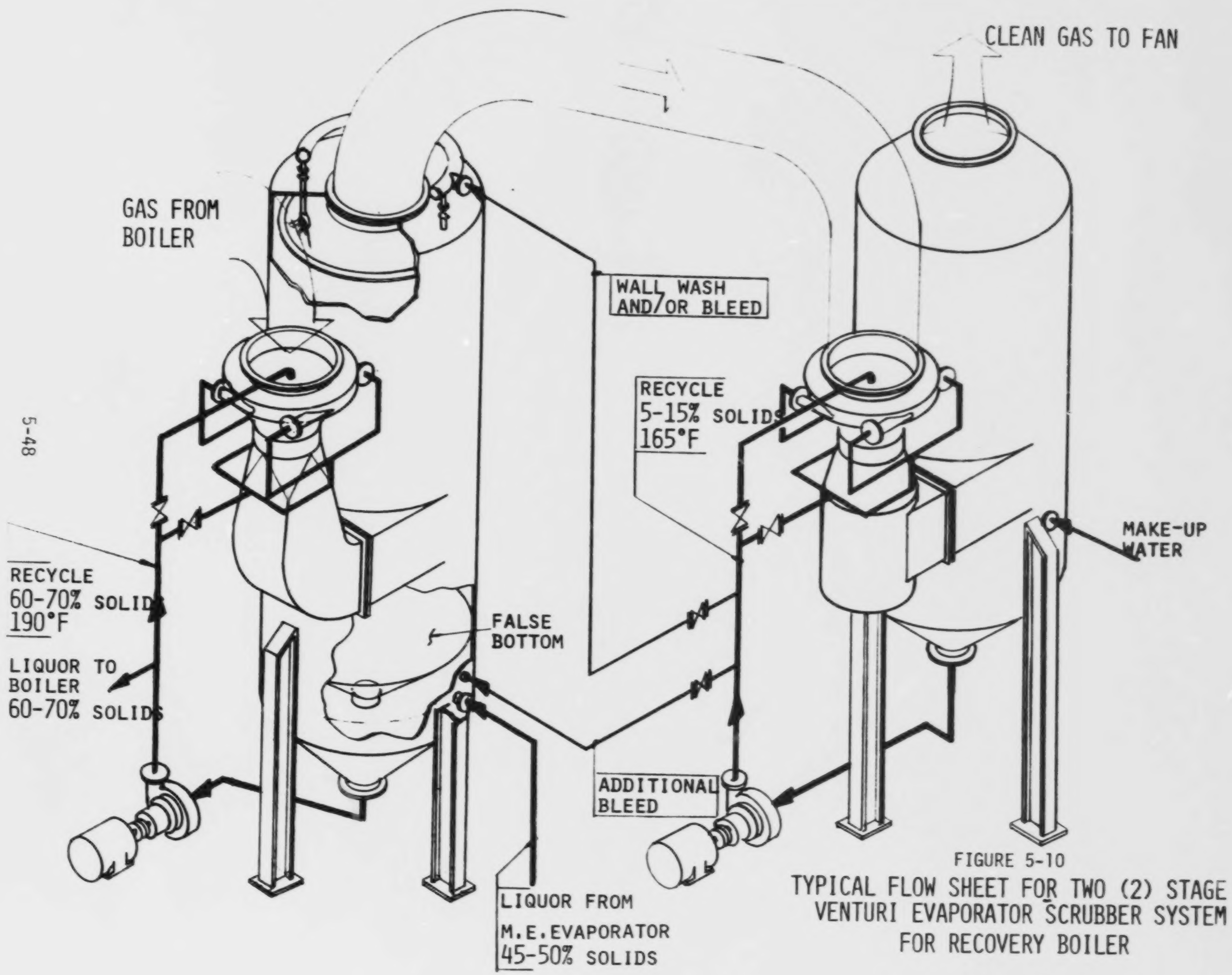

CLEAN GAS TO FAN
GAS FROM BOILER
WALL WASH AND/OR BLEED
RECYCLE 5-15% SOLIDS 165°F
MAKE-UP WATER
RECYCLE 60-70% SOLIDS 190°F
FALSE BOTTOM
LIQUOR TO BOILER 60-70% SOLIDS
ADDITIONAL BLEED
LIQUOR FROM M.E.EVAPORATOR 45-50% SOLIDS

FIGURE 5-10
TYPICAL FLOW SHEET FOR TWO (2) STAGE VENTURI EVAPORATOR SCRUBBER SYSTEM FOR RECOVERY BOILER

is rather small, it does not affect the overall heat balance, but helps to reduce the steam plume. No wall wash is necessary in the separator of the second stage.

This type of Venturi, used in both the first and the second stage, does not have any nozzles, weirs or narrow restrictions, which can plug up or cause dust build-ups. No steam atomization is, therefore, necessary. Since the solids concentration in the recycle liquid of the second stage is low, a higher collection efficiency can be attained for the same total pressure drop as compared with the one-stage evaporator-scrubber. The two-stage system also has a higher thermal efficiency than the single stage system.

For both types of systems, the scrubbers have to be corrosion resistant.

The fan is normally located between the scrubber and the stack. Due to the corrosive atmosphere, the fan has to be constructed in stainless steel, be provided with split housing for easy wheel removal, and have inspection doors, water sprays and drain connections. Ductwork is normally made from stainless steel.

The recirculation pump is normally made from corrosion resistant material.

Recovery Furnace Scrubbing Liquors. The liquid scrubbing systems which have been described depend to a considerable extent for their effectiveness on selection of a suitable liquid phase. In some instances, the scrubber may remove both particulates and gaseous compounds from the gas stream. The use of some liquids may actually add to the concentration of odorous compounds in the flue gases.

Selection of a scrubbing liquor for removing particulate matter and odoriferous gas components from recovery boiler flue gases is a difficult task because the gas volume is large and the CO_2 content is very high resulting in the formation of H_2S in the presence of Na_2S as follows:

$$H_2O + CO_2 \rightleftarrows H_2CO_3$$

$$H_2CO_3 + Na_2S \rightleftarrows Na_2CO_3 + H_2S$$

Therefore, if either weak wash from the causticizing operation or weak black liquor is used as the scrubbing liquor, it must first be oxidized to change the Na_2S content to $Na_2S_2O_3$. However, Harding and Galeano ((7) have shown that above an oxidation efficiency of 80 percent, the absorption of odoriferous gases decreases because the NaOH originally formed by the oxidation reaction is destroyed by further reaction with O_2 and $Na_2S_2O_3$ as shown below:

$$2Na_2S + 2O_2 + H_2O \rightarrow Na_2S_2O_3 + 2NaOH$$

$$Na_2S_2O_3 + 2NaOH + 2O_2 \rightarrow 2Na_2SO_4 + H_2O$$

After studying the above reactions, the successful use of Na_2S containing liquors is doubtful since the presence of Na_2S in partially oxidized liquor results in formation of additional H_2S, while the use of 99 percent plus oxidized liquor results in greatly reduced H_2S absorption efficiency.

Water has been moderately successful in removing relatively large particulate matter in low pressure drop scrubbers following electrostatic precipitators. More efficient particulate removal can be obtained with water if high pressure drop Venturis are used. However, the efficiency of water in scrubbing odoriferous gases is very low regardless of the equipment used.

In summary, technology has yet to produce a tried and proven system for simultaneously eliminating both gases and particulates from recovery furnace flue gases with any of the scrubbing liquors available for use. More research pilot plant data and operating experience with systems using different types of scrubbing liquors must be done before any valid conclusions can be reached in the selection of the most feasible scrubbing liquor.

Case Descriptions

The following cases have been considered for application of control methods to recovery systems.

Case 1. Replacement of an existing 90 percent AOE precipitator with a higher efficiency precipitator.

Case 2. Installation of additional control equipment in series with an existing 90 percent AOE precipitator.

Case 3. Installation of a cyclonic scrubber in series with an existing 95 percent precipitator located on the roof.

Case 4. Application of additional control methods to an existing recovery system with an 80 percent AOE black liquor Venturi.

Case 5. Installation of a black liquor oxidation system to an existing recovery system.

Case 6. Capital Costs of new recovery systems.

5.3.6.1.1 Recovery System, Case 1

Application

This case is based on the replacement of an existing 90 percent AOE precipitator with a higher efficiency precipitator. Most of the existing older recovery precipitator installations are arranged with the precipitator on the ground. Usually there is no additional space on the ground in the vicinity of the recovery unit; therefore, an additional precipitator would have to be installed above the existing precipitator. Further, the existing precipitator could not be removed and replaced with a new one since this would require a lengthy mill shutdown resulting in a high dollar cost due to the loss of production.

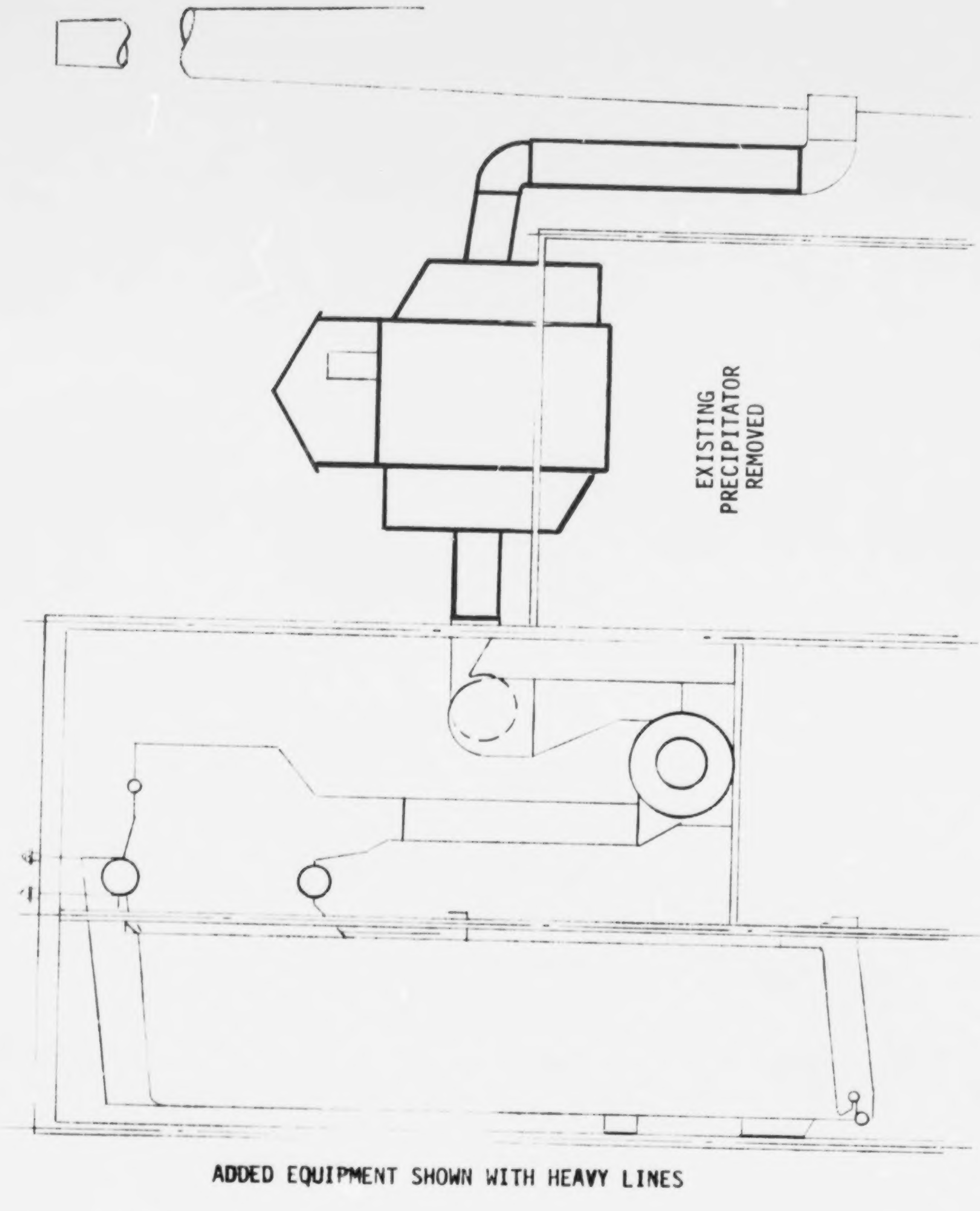
Costs

The capital costs are based on a couble chamber, common wall, tile construction precipitator arranged as shown in Figure 5-11. Auxiliaries included are: agitators, dampers, circulation pumps, instruments and controls for proper operation of the above, and revised ductwork to connect the addition of replacement precipitator in the system. The cost also reflects the structure required to support this addition, and the demolition cost for removal of the existing precipitator. Capital costs and net annual costs are presented in Figures 5-12, 5-13, and 5-14.

Effectiveness

Particulate removal. Methods are considered for replacing a 90 percent AOE precipitator on an existing recovery boiler. These methods and their particulate efficiencies are as follows:

Control Method	Guaranteed Efficiency	Annual Operating Efficiency
Precipitator	99.9	99.5
Precipitator	99.5	99.0
Precipitator	99.0	98.5


EXISTING PRECIPITATOR REMOVED

ADDED EQUIPMENT SHOWN WITH HEAVY LINES

FIG. 5 - 11 TYPICAL ARRANGEMENT FOR ELECTROSTATIC PRECIPITATOR REPLACING AN EXISTING PRECIPITATOR - RECOVERY BOILER

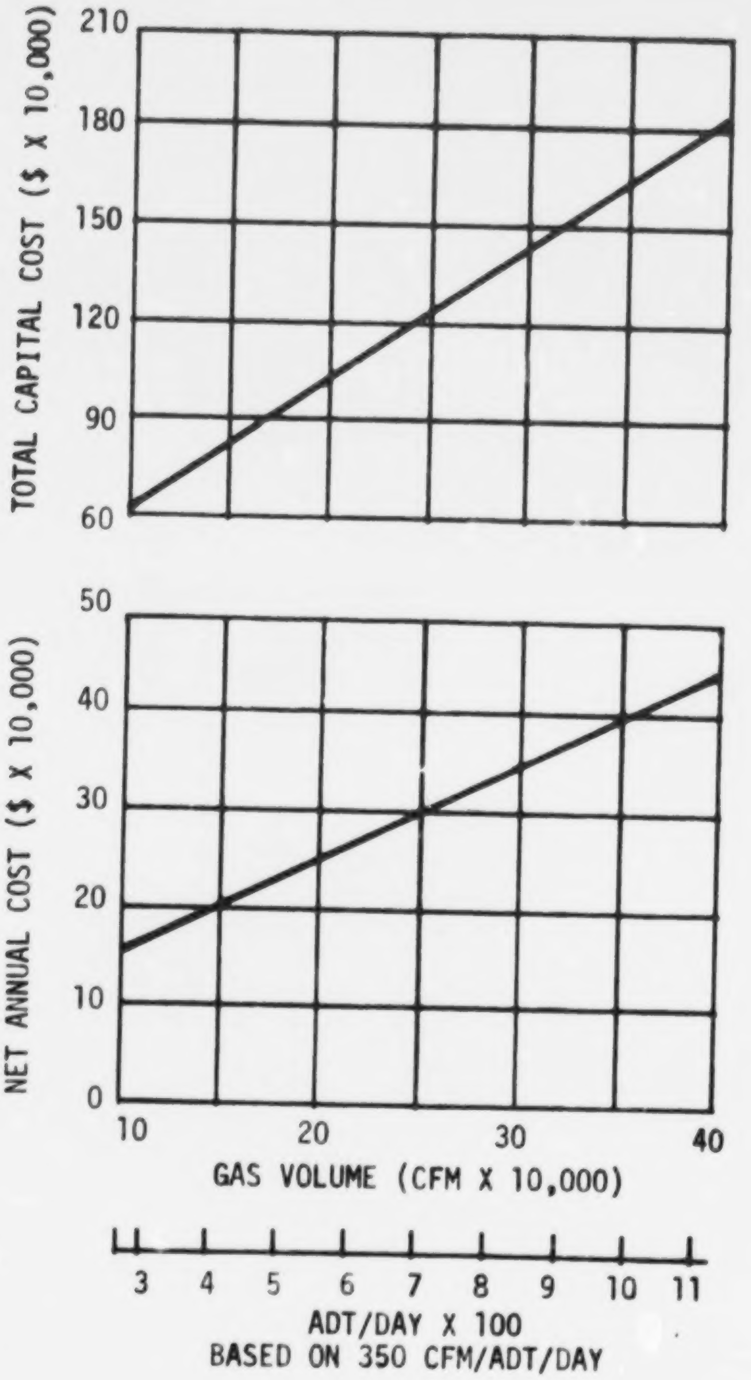


FIG. 5 - 12 CONTROL METHOD COSTS FOR 99.9% EFF. ELECTROSTATIC PRECIPITATOR REPLACING AN EXISTING PRECIPITATOR - RECOVERY BOILER (99.5% A.O.E.)

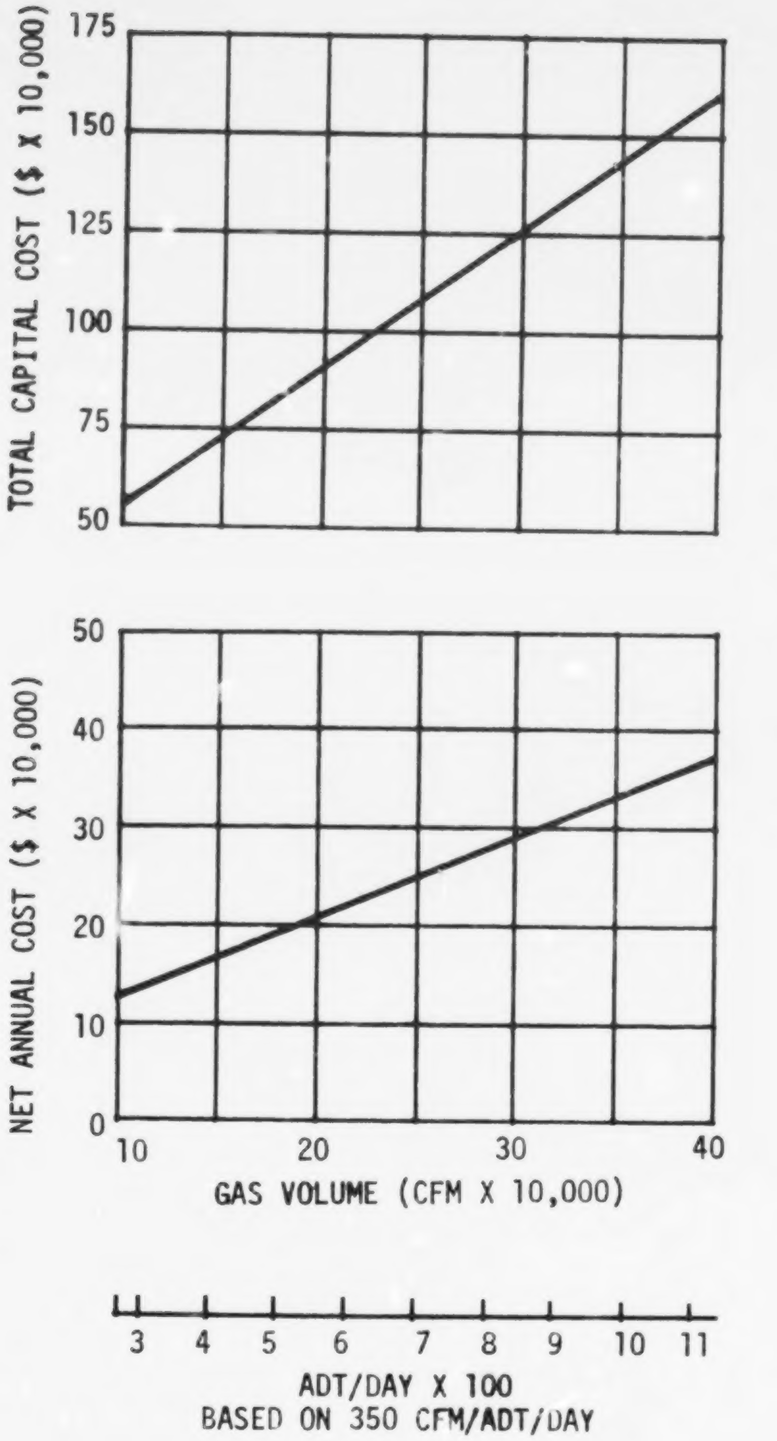


FIG. 5 - 13 CONTROL METHOD COSTS FOR 99.5% EFF. ELECTROSTATIC PRECIPITATOR REPLACING AN EXISTING PRECIPITATOR - RECOVERY BOILER (99.0% A.O.E.)

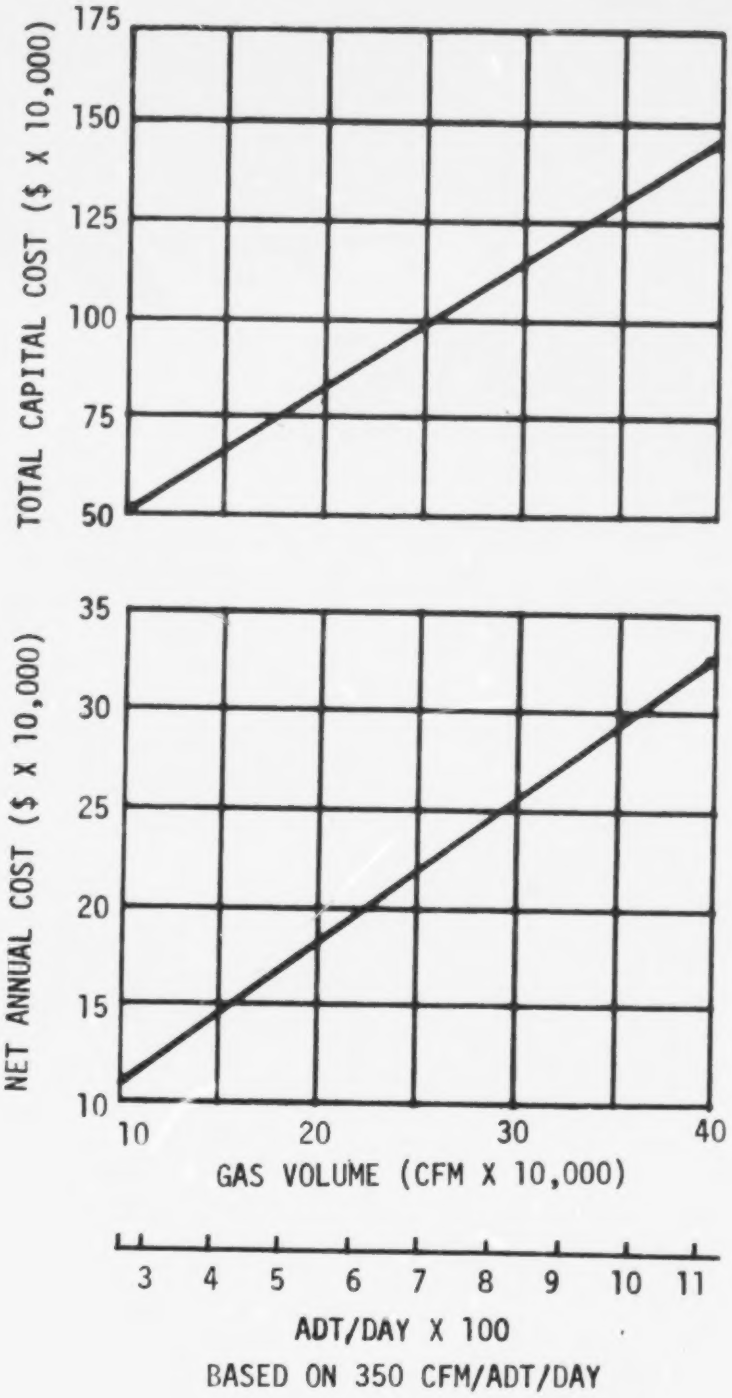
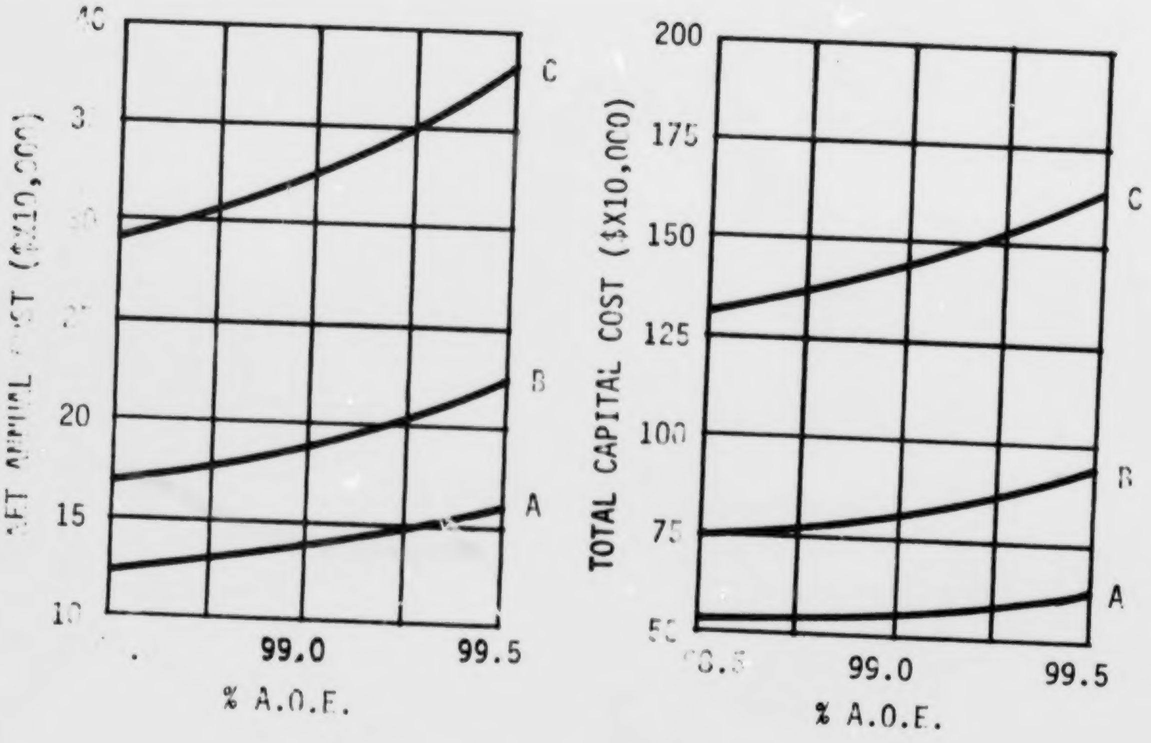


FIG. 5 - 14 CONTROL METHOD COSTS FOR 99.0% EFF. ELECTROSTATIC PRECIPITATOR REPLACING AN EXISTING PRECIPITATOR - RECOVERY BOILER (98.5% A.O.E.)

The attached cost effectiveness curves (Figure 5-15) indicate the change in costs as efficiency and size of equipment are varied.

Reduced Sulfur and Sulfur Dioxide Removal. Since precipitators do not remove either reduced sulfur or sulfur oxides, all three methods would be equally ineffective on such emissions.

Operation. All three control methods should be equal in terms of operation.



A. 105,000 CFM, (300 ADT/DAY)
B. 175,000 CFM, (500 ADT/DAY)
C. 350,000 CFM, (1000 ADT/DAY)

NOTE - ADT/DAY BASED ON 350 CFM/ADT/DAY

FIG. 5-15 PARTICULATE COST EFFECTIVENESS FOR ELECTROSTATIC PRECIPITATOR REPLACING AN EXISTING PRECIPITATOR - RECOVERY BOILER

5.3.6.1.2 Recovery System, Case 2

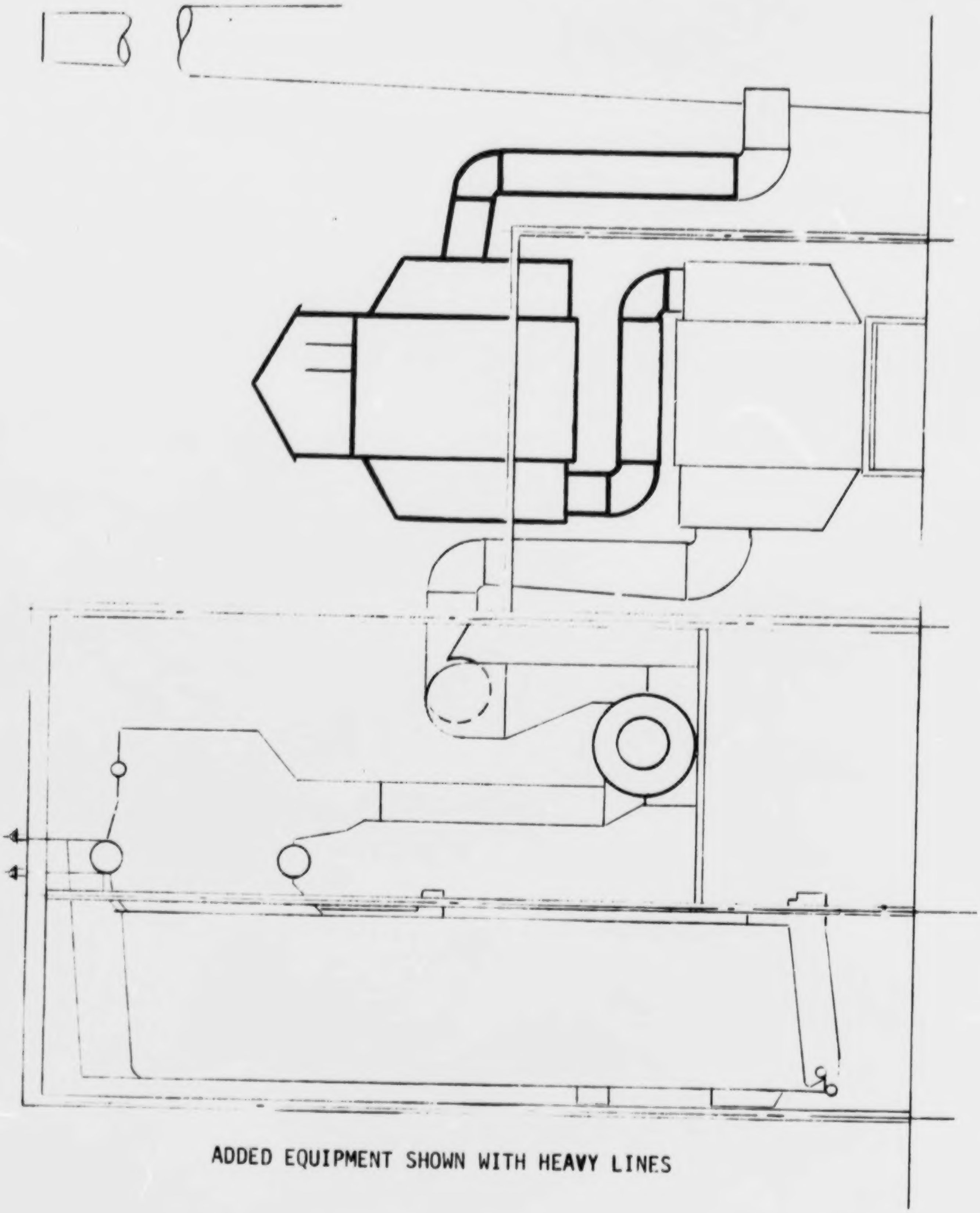
Application

This case is based on the installation of additional control equipment in series with an existing 90 percent AOE precipitator located on the ground. The control methods considered are electrostatic precipitators, Venturi scrubbers, and cyclone scrubbers.

Costs

Electrostatic Precipitator. This control method is based on the installation of an electrostatic precipitator in series with and located above an existing precipitator as indicated in Figure 5-16. This location was chosen as the likelihood of having room for adding a second precipitator on the ground would be practically nil in an average mill. The capital cost and net annual cost shown in Figures 5-17, 5-18, and 5-19 are based on a double chamber, common wall, tile construction precipitator with a wet bottom and include the auxiliaries such as agitators, dampers, circulation pumps, instruments and controls for proper operation of the above, and revised ductwork to connect the additional precipitator to the existing precipitator and to the existing stack. The costs also reflect the structure required to support this addition.

Venturi scrubber. This control method is based on the installation of a 304 stainless steel Venturi and a concrete lined mild steel separator in series with an existing electrostatic precipitator as indicated in Figure 5-20. The outlet duct and additional I.D. fan are 304 stainless steel construction. Due to the addition of the Venturi, the pressure drop through the exhaust gas system will be greatly increased requiring additional fan pressure. This case is based on an additional variable speed I.D. fan being installed between the Venturi and the existing stack. Capital costs and net annual costs are presented in Figures 5-21, 5-22, and 5-23.



ADDED EQUIPMENT SHOWN WITH HEAVY LINES

FIG. 5 - 16 TYPICAL ARRANGEMENT FOR ELECTROSTATIC PRECIPITATOR ADDED IN SERIES WITH EXISTING PRECIPITATOR - RECOVERY BOILER

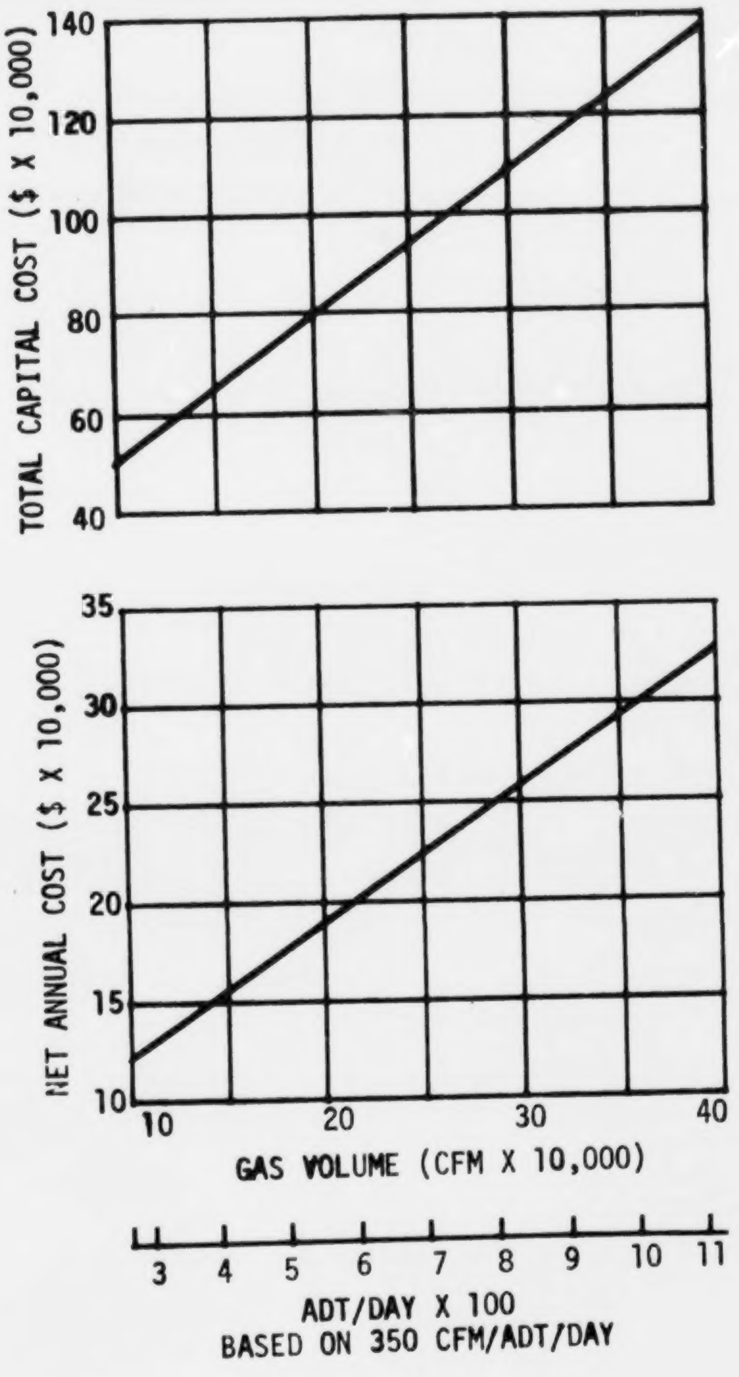


FIG. 5 - 17 CONTROL METHOD COSTS FOR 99% EFF. ELECTROSTATIC PRECIPITATOR ADDED IN SERIES WITH AN EXISTING PRECIPITATOR-RECOVERY BOILER (99.8% A.O.E.)

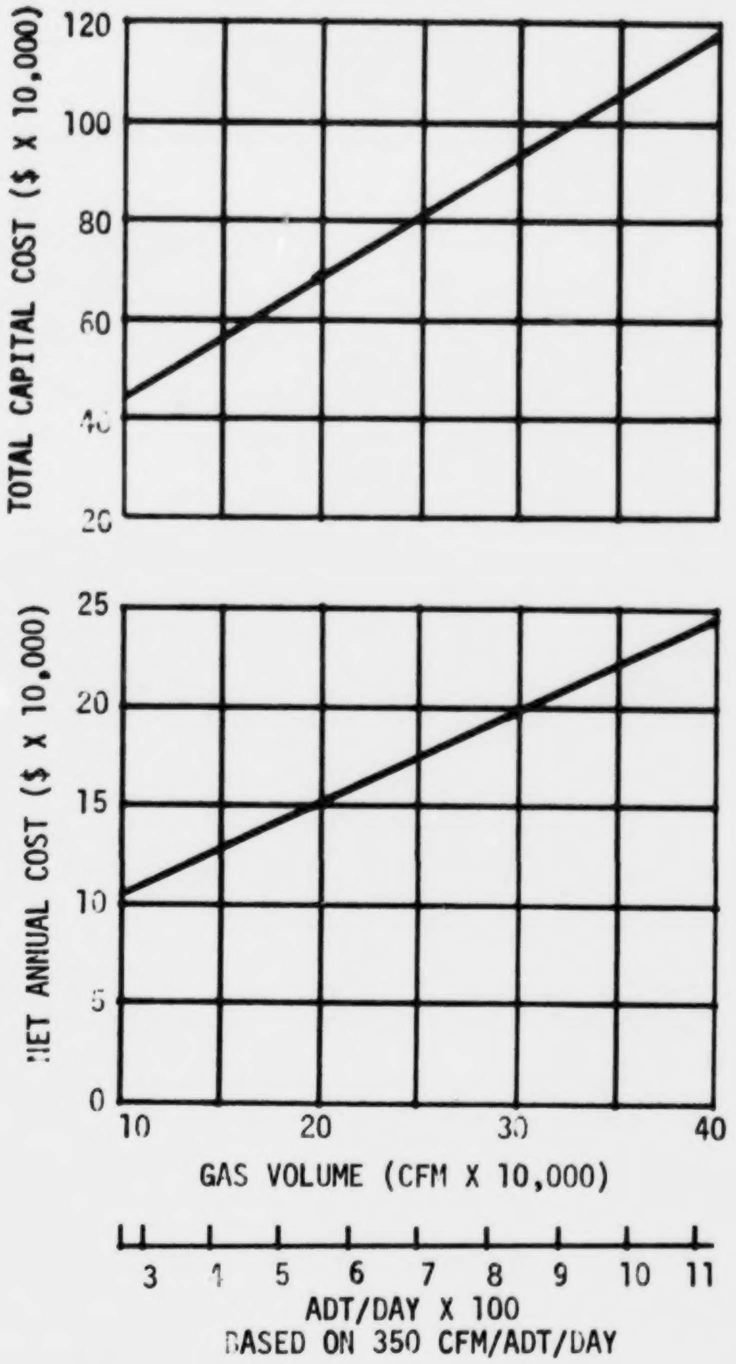


FIG. 5 - 18 CONTROL METHOD COST FOR 95% EFFICIENCY ELECTROSTATIC PRECIPITATOR ADDED IN SERIES WITH AN EXISTING PRECIPITATOR-RECOVERY BOILER (99.4% A.O.E.)

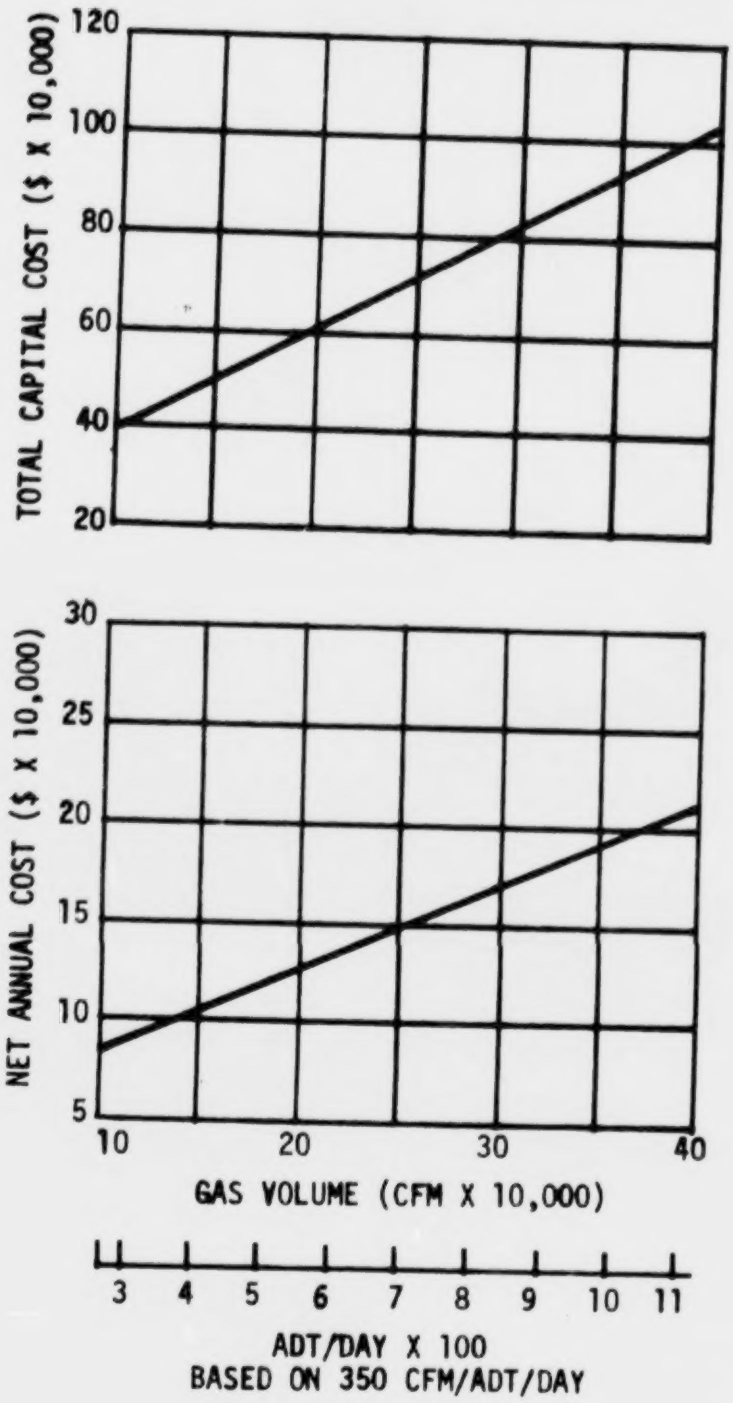
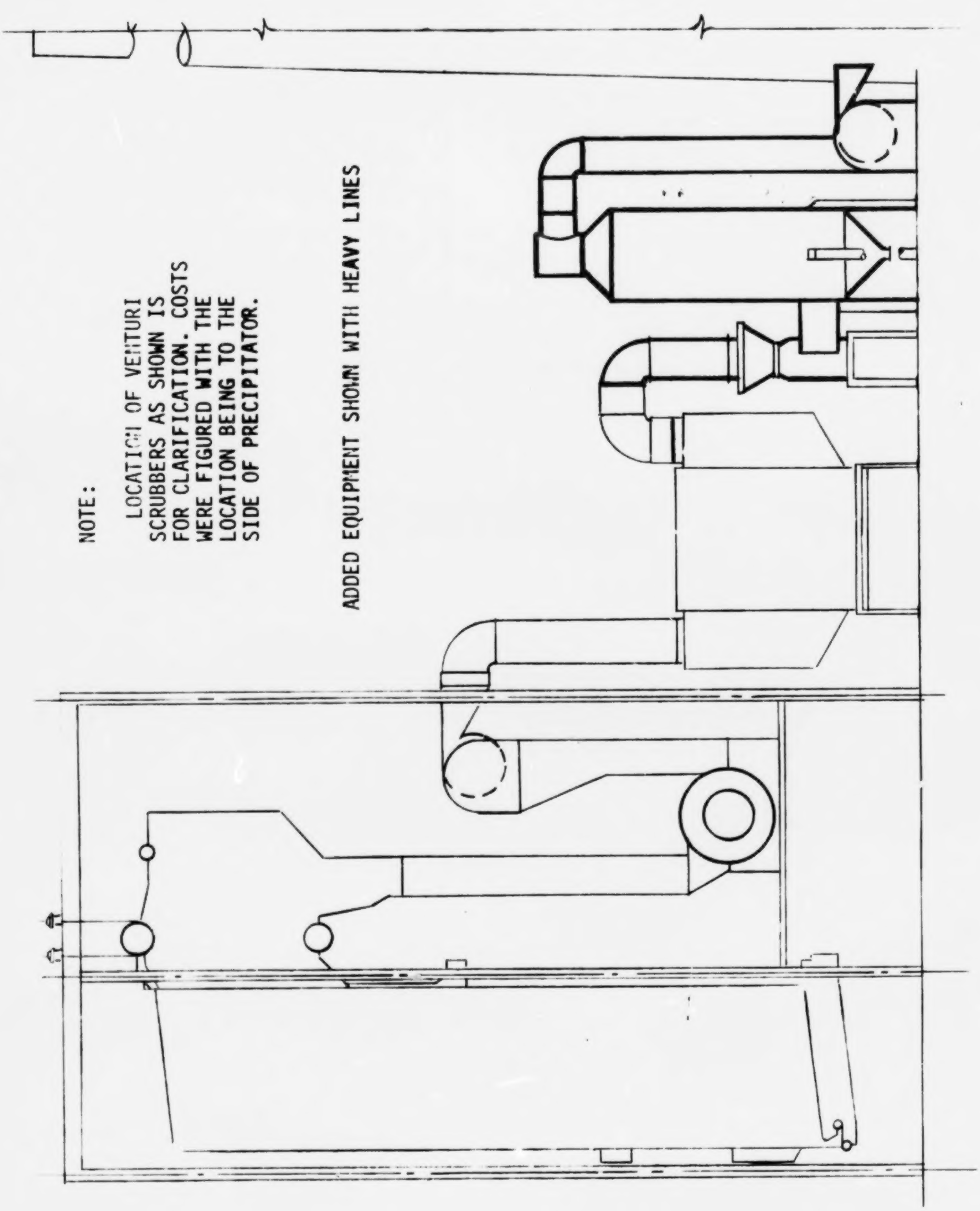


FIG. 5 - 19 CONTROL METHOD COSTS FOR 90% EFF. ELECTROSTATIC PRECIPITATOR ADDED IN SERIES WITH AN EXISTING PRECIPITATOR-RECOVERY BOILER (98.9% A.O.E.)



NOTE:

LOCATION OF VENTURI SCRUBBERS AS SHOWN IS FOR CLARIFICATION. COSTS WERE FIGURED WITH THE LOCATION BEING TO THE SIDE OF PRECIPITATOR.

ADDED EQUIPMENT SHOWN WITH HEAVY LINES

FIG. 5 - 20 TYPICAL ARRANGEMENT FOR VENTURI SCRUBBER ADDED TO AN EXISTING PRECIPITATOR - RECOVERY BOILER

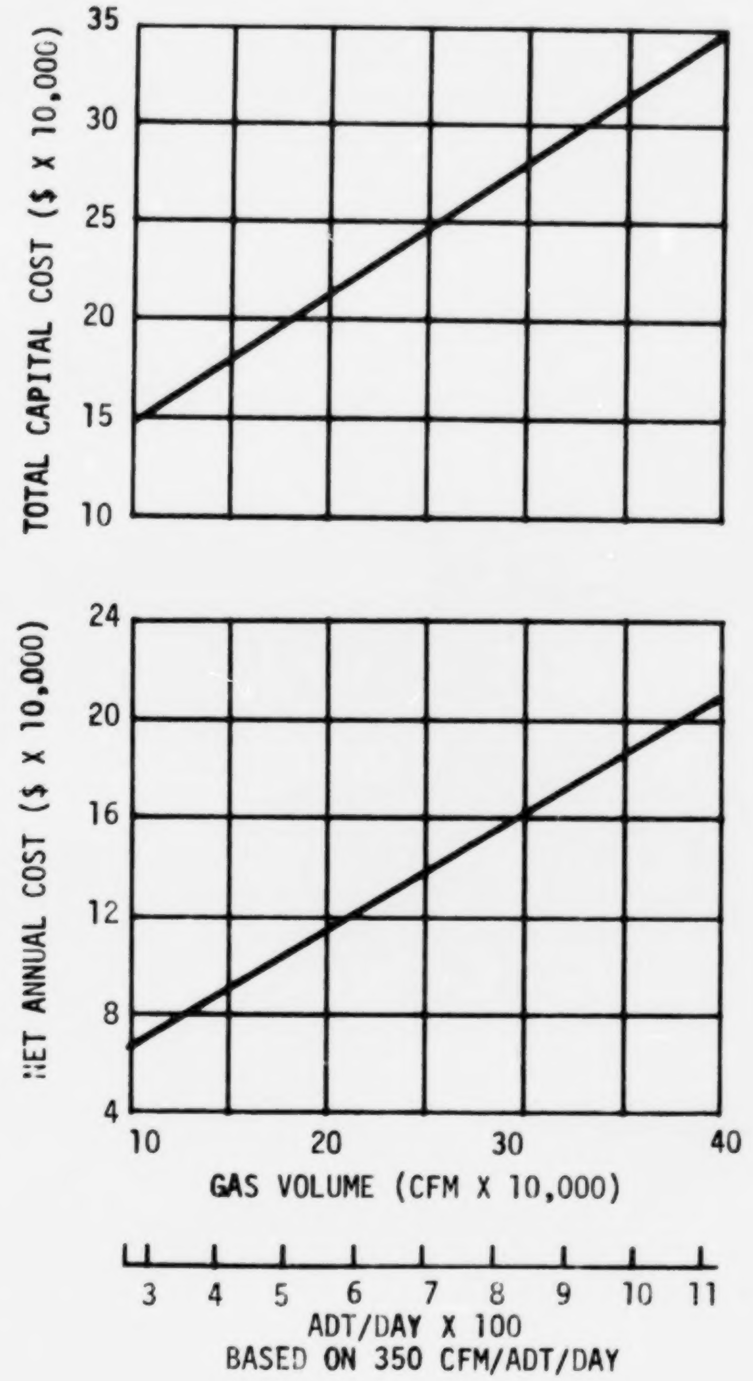


FIG. 5 - 21 CONTROL METHOD COSTS FOR A 99% EFF. VENTURI SCRUBBER ADDED TO AN EXISTING 90% EFF. PRECIPITATOR - RECOVERY BOILER (99.8% A.O.E.)

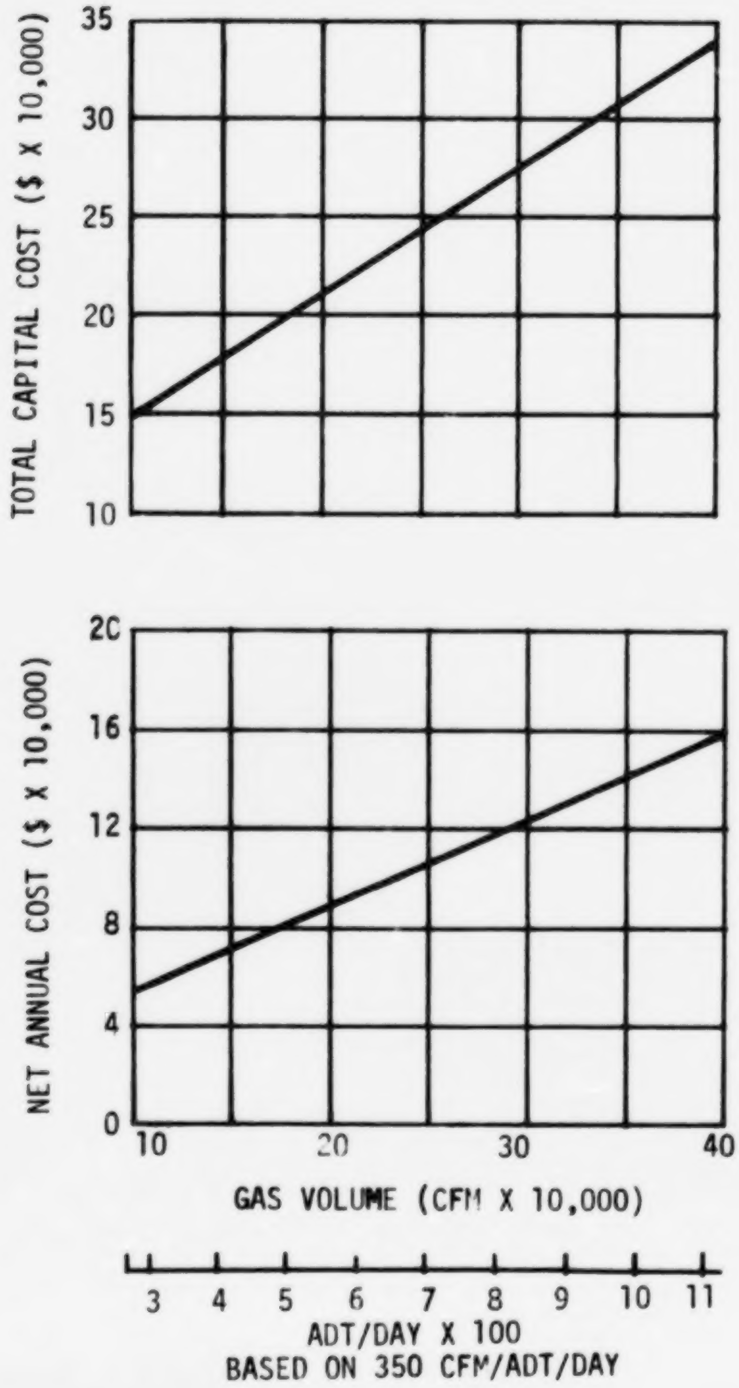


FIG. 5 - 22 CONTROL METHOD COSTS FOR A 95% EFF. VENTURI SCRUBBER ADDED TO AN EXISTING 90% EFF. PRECIPITATOR - RECOVERY BOILER (99.4% A.O.E.)

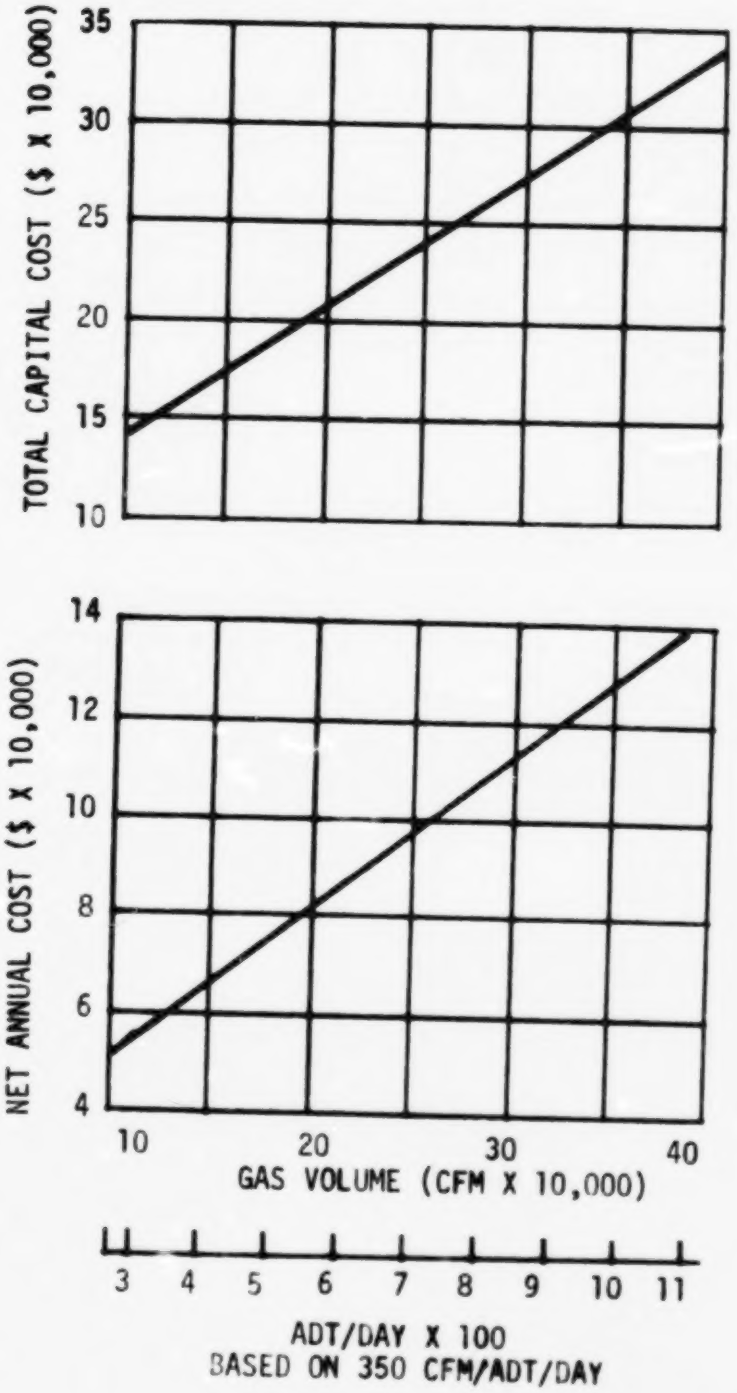


FIG. 5 - 23 CONTROL METHOD COSTS FOR A 90% EFF. VENTURI SCRUBBER ADDED TO AN EXISTING 90% EFF. PRECIPITATOR-RECOVERY BOILER (98.9% A.O.E.)

In addition to the Venturi and fan, the costs include a recirculattion pump and motor, instruments and controls for proper operation, required piping for make-up, recirculation pump and motor, instruments and controls for proper operation, required piping for make-up, recirculation and effluent, and revised ductwork to connect the Venturi with the existing precipitator and stack.

Gas volumes are actual duct conditions at the inlet to the Venturi. Particulate removal efficiencies are based on a loading of 1.8 grains per actual cubic foot.

Cyclonic scrubber. This control method is based on the installation of two or more cyclonic scrubbers in series with an existing electrostatic precipitator as shown in Figure 5-24. In addition to the scrubbers, a recirculation pump and motor for each scrubber, instruments and controls for proper operation of the above, required piping for make-up, recirculation, and effluent to the precipitator, and revised ductwork to connect the scrubbers with the existing system are included. One 20-foot stub stack is included for each scrubber located on the roof, while additional duct is included to connect each scrubber located on the ground to the existing stack.

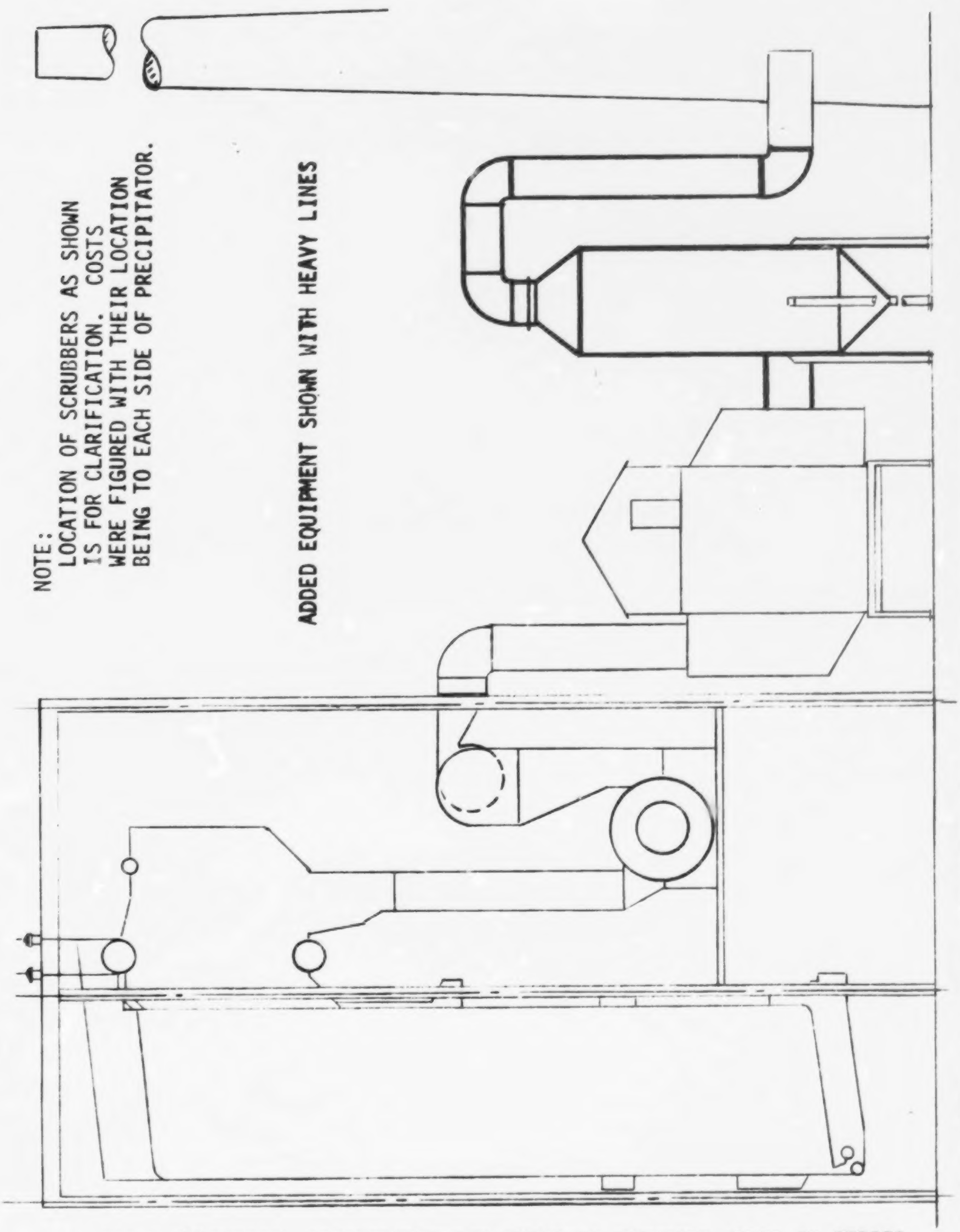
There are several widely used materials of construction for cyclonic scrubbers, 304 and 316 stainless steel, fiberglass, and mild steel with concrete lining. The capital cost is based on 304 stainless steel cyclone and outlet duct or stack as this would be the most representative material for this application; inlet duct is carbon steel.

Gas volumes are actual duct conditions at the inlet to the scrubber. Particulate removal efficiencies are based on a loading of 1.8 grains per actual cubic foot.

Capital costs and annual operating costs are presented in Figure 5-25.

Effectiveness

Particulate removal. Considerations for this case are based on an existing recovery furnace operating with a 90 percent AOE precipitator. Various control methods are considered for increasing the particulate efficiency to levels of 98.0, 99.4, and 99.8 percent AOE. These control methods are as follows:



NOTE:
LOCATION OF SCRUBBERS AS SHOWN IS FOR CLARIFICATION. COSTS WERE FIGURED WITH THEIR LOCATION BEING TO EACH SIDE OF PRECIPITATOR.

ADDED EQUIPMENT SHOWN WITH HEAVY LINES

FIG. 5-24 TYPICAL ARRANGEMENT FOR CYCLONIC SCRUBBER ADDED IN SERIES WITH EXISTING PRECIPITATOR ON GROUND - RECOVERY BOILER

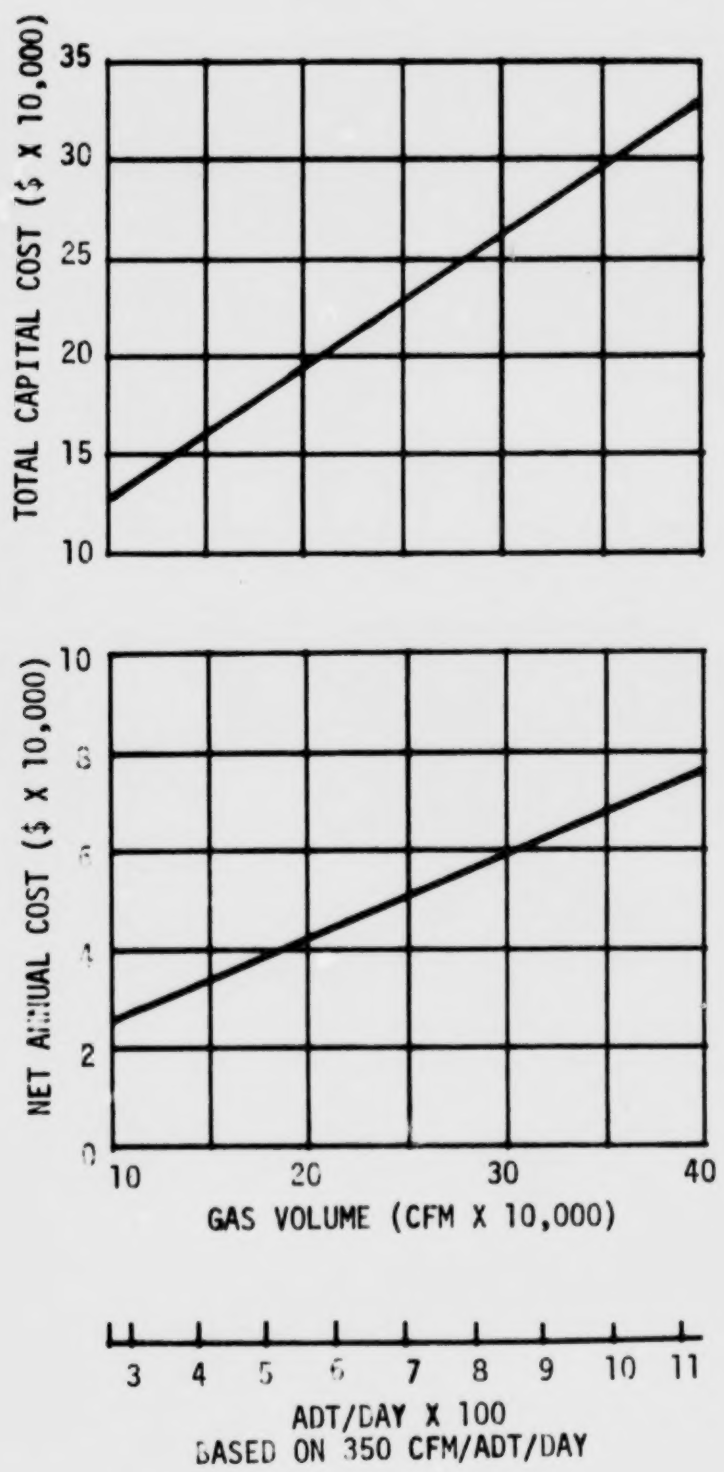


FIG. 5 - 25 CONTROL METHOD COSTS FOR CYCLONIC SCRUBBER ADDED IN SERIES WITH EXISTING PRECIPITATOR ON GROUND - RECOVERY BOILER

Control Method	Guaranteed Efficiency	AOE of New Equip.	AOE of Exist. Equip.	Total AOE
Precipitator	99.0	98.5	90.0	99.8
Venturi Scrubber	99.0	98.0	90.0	99.8
Precipitator	95.0	94.5	90.0	99.4
Venturi Scrubber	95.0	94.0	90.0	99.4
Precipitator	90.0	89.5	90.0	98.9
Venturi Scrubber	90.0	89.0	90.0	98.9
Cyclone Scrubber		80.0	90.0	98.0

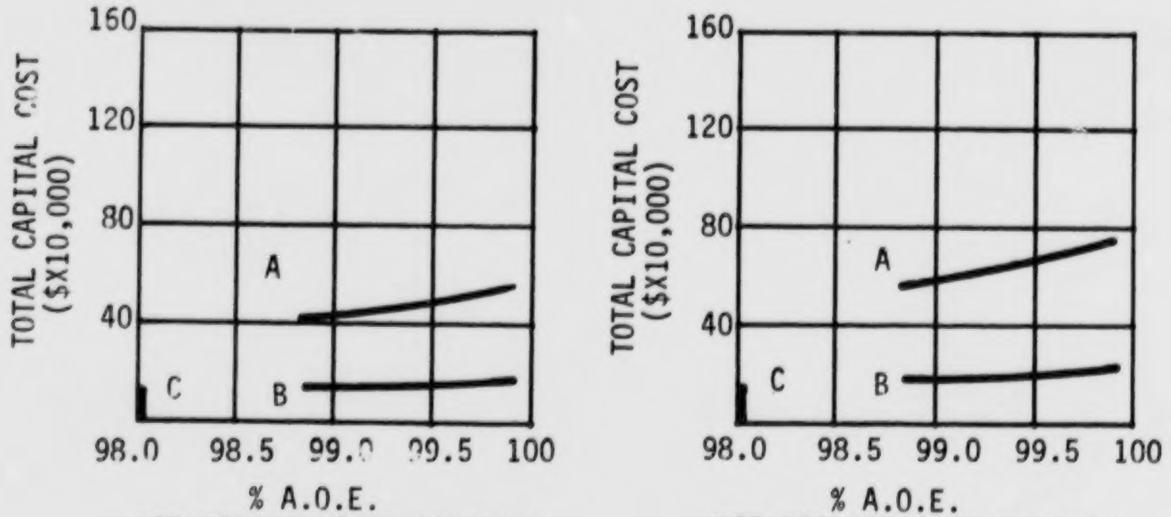
Curves for particulate cost effectiveness are presented in Figures 5-26 and 5-27.

Reduced Sulfur and Sulfur Dioxide Removal. Precipitators remove particulate only and cannot absorb either reduced sulfur or sulfur oxides. Depending on the scrubbing liquid, both the Venturi and cyclonic scrubbers have the potential for absorbing reduced sulfur and sulfur oxides. However, present technology is not well enough developed to assign values for absorption efficiencies.

Scrubbers may also increase the sulfur compounds in the flue gas with certain scrubbing liquids (e.g., unoxidized black liquor).

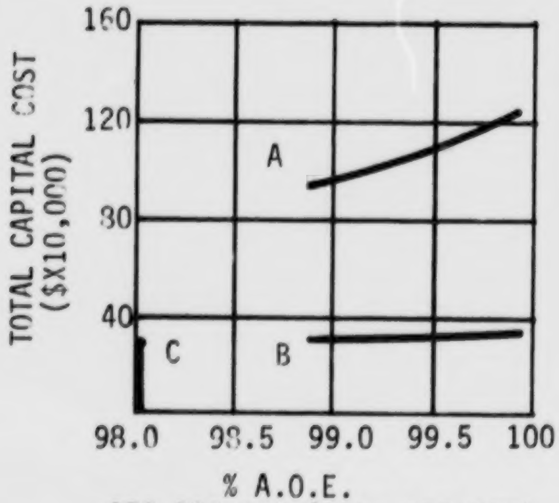
Operation. As compared to precipitators, Venturi or cyclonic scrubbers produce a lower temperature, higher moisture content gas. These conditions can result in more corrosion and a lower plume rise.

Summary. The cyclonic scrubber has the lowest cost of the methods considered; however, the efficiency is limited to 98.0 percent. Where higher efficiencies are required, the Venturi should be the most effective method.



105,000 CFM,(300 ADT/DAY)

175,000 CFM,(500 ADT/DAY)

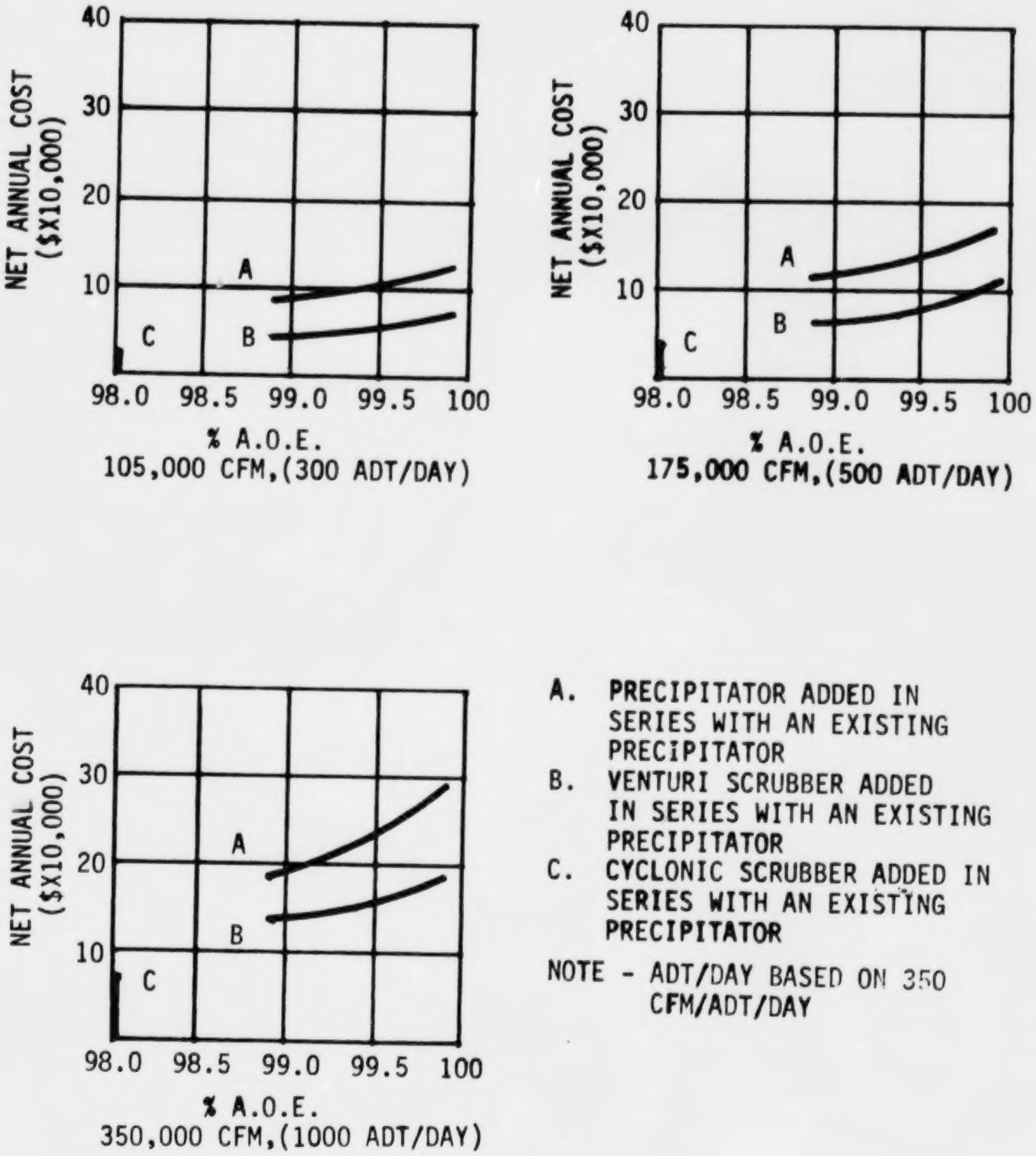


350,000 CFM,(1000 ADT/DAY)

A. PRECIPITATOR ADDED IN SERIES WITH AN EXISTING PRECIPITATOR
B. VENTURI SCRUBBER ADDED IN SERIES WITH AN EXISTING PRECIPITATOR
C. CYCLONIC SCRUBBER ADDED IN SERIES WITH AN EXISTING PRECIPITATOR

NOTE - ADT/DAY BASED ON 350 CFM/ADT/DAY

FIG.5-26 PARTICULATE COST EFFECTIVENESS (TOTAL CAPITAL COST COMPARISON) FOR ADD-ON CONTROL EQUIPMENT - RECOVERY BOILER



105,000 CFM,(300 ADT/DAY)

175,000 CFM,(500 ADT/DAY)

350,000 CFM,(1000 ADT/DAY)

A. PRECIPITATOR ADDED IN SERIES WITH AN EXISTING PRECIPITATOR

B. VENTURI SCRUBBER ADDED IN SERIES WITH AN EXISTING PRECIPITATOR

C. CYCLONIC SCRUBBER ADDED IN SERIES WITH AN EXISTING PRECIPITATOR

NOTE - ADT/DAY BASED ON 350 CFM/ADT/DAY

FIG.5 - 27 PARTICULATE COST EFFECTIVENESS (NET ANNUAL COST COMPARISON) FOR ADD-ON CONTROL EQUIPMENT - RECOVERY BOILER

5.3.6.1.3 Recovery System, Case 3

Application

This case is based on the installation of cyclonic scrubbers on the roof of an existing recovery building. The scrubbers would be in series with and following an existing 95 percent AOE precipitator also located on the roof.

Costs

Estimated costs are based on the equipment arrangement shown in Figure 5-28. These costs are shown in Figure 5-29 and include the scrubbers, a recirculation pump and motor for each scrubber, instruments and controls, piping for make-up, recirculation, and effluent to the precipitator, revised ductwork to connect the scrubbers with the existing system, and one 20-foot stack for each scrubber. It was assumed that allowances were made in the original building design for future scrubbers on the roof, and no costs for building structure are included.

There are several widely used materials of construction for cyclonic scrubbers, 304 and 316 stainless steel, fiberglass, and mild steel with concrete lining. The capital cost is based on 304 stainless steel cyclone and outlet duct or stack as this would be the most representative material for this application; inlet duct is carbon steel.

Gas volumes are actual duct conditions at the inlet to the scrubber. Particulate removal efficiencies are based on a loading of 1.8 grains per actual cubic foot.

Effectiveness

Particulate Removal. The addition of cyclonic scrubbers to an existing 95 percent AOE precipitator would result in an expected overall AOE of 99 percent.

Reduced Sulfur and Sulfur Dioxide Removal. Cyclonic scrubbers have the potential for either removing or releasing sulfur compounds depending on the scrubbing liquid utilized.

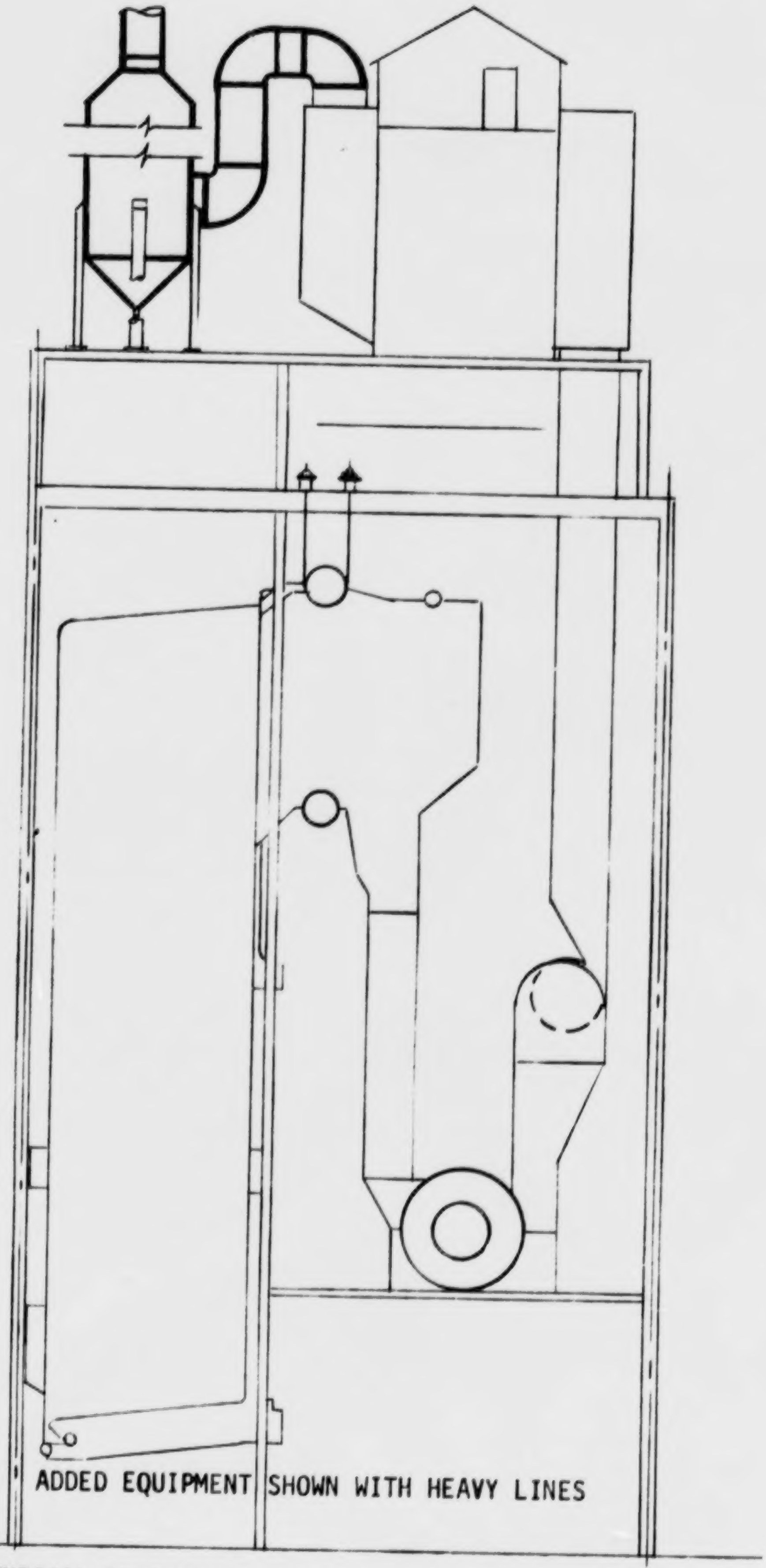

ADDED EQUIPMENT SHOWN WITH HEAVY LINES

FIG.5-28 TYPICAL ARRANGEMENT FOR CYCLONIC SCRUBBER ADDED IN SERIES WITH EXISTING PRECIPITATOR ON ROOF - RECOVERY BOILER

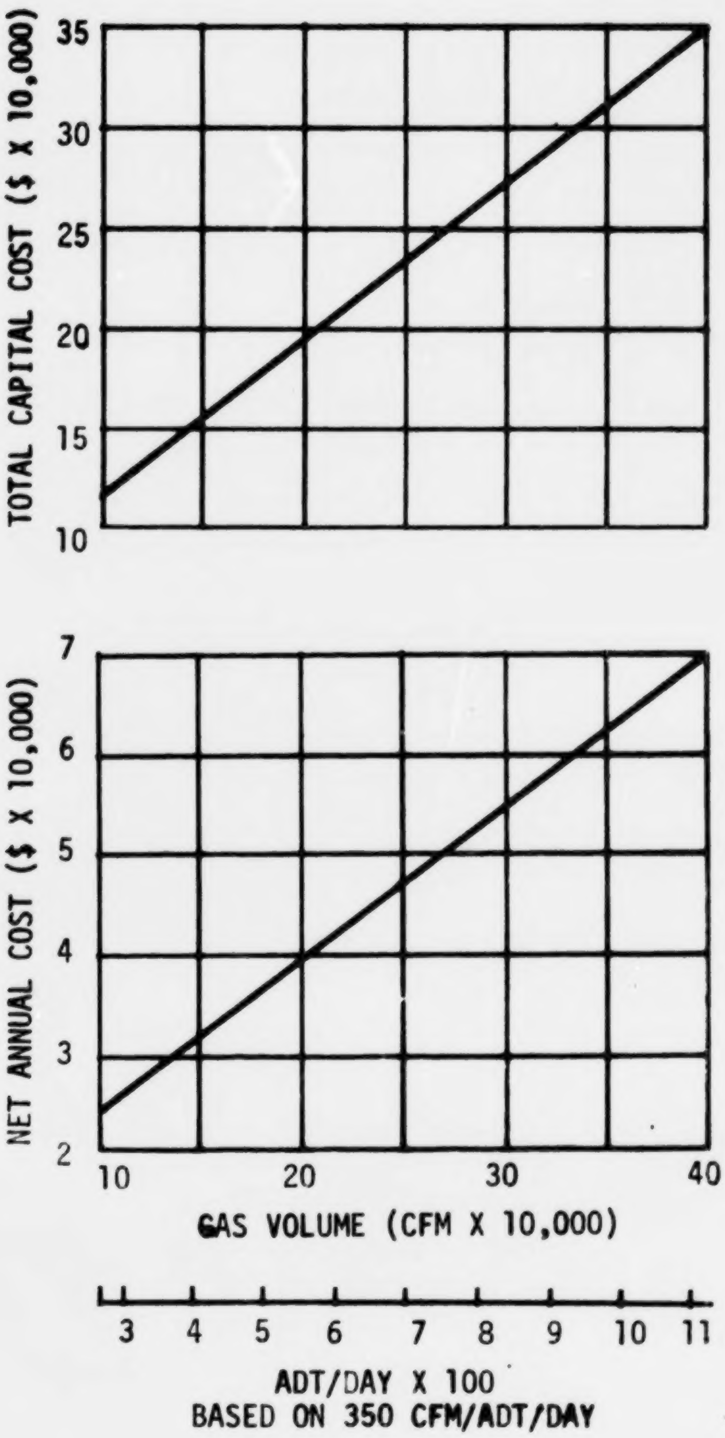


FIG. 5 - 29 CONTROL METHOD COSTS FOR CYCLONIC SCRUBBER ADDED IN SERIES WITH EXISTING PRECIPITATOR ON ROOF - RECOVERY BOILER

Operation. Possible operating disadvantages of the cyclonic scrubbers would be corrosion potential and lower plume rise resulting from high moisture.

Summary. The addition of cyclonic scrubbers appears to be the only practical method of increasing the efficiency of an existing precipitator located on the roof. The possibility of installing Venturi scrubbers was not considered in detail because space is not usually available for the required addition of an I. D. fan and ductwork.

5.3.6.1.4 Recovery System, Case 4

Application

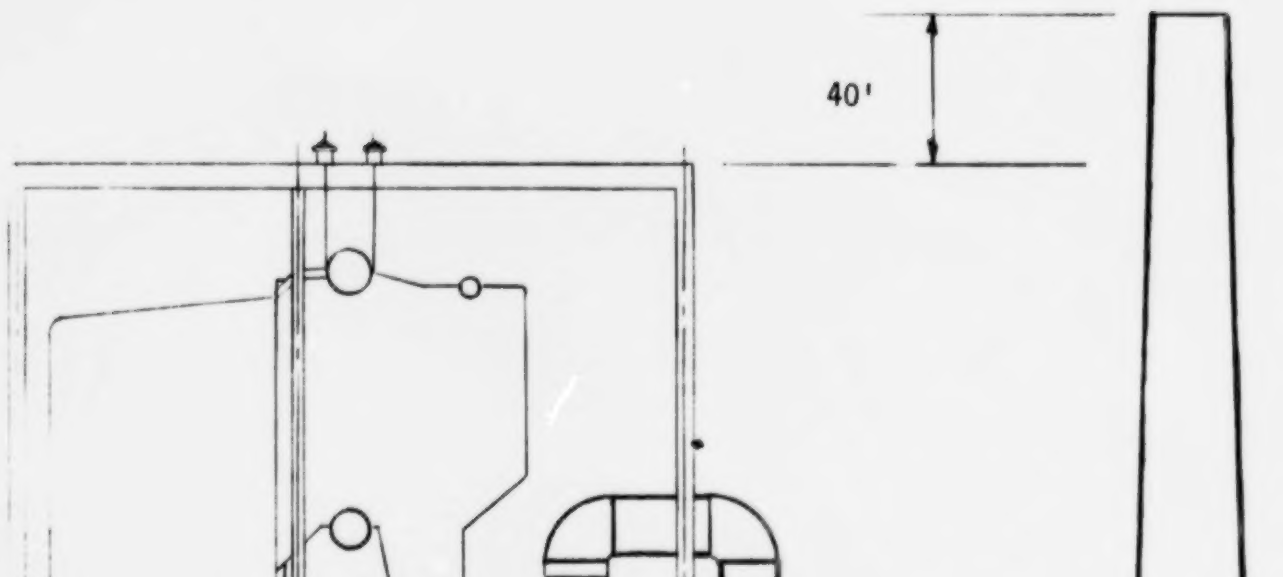
This case is based on applying additional control methods to an existing recovery system with an 80 percent AOE black liquor venturi. The methods considered are as follows:

1. Convert Venturi to cyclone evaporator and add an electrostatic precipitator.
2. Convert existing Venturi to low energy type and add a second stage Venturi scrubber.

Costs

Precipitator. Control method costs are based on the equipment arrangement shown in Figure 5-30. These costs as shown in Figures 5-31 and 5-32 include the addition of a double chamber, common wall, tile construction precipitator and the removal of the Venturi section and the addition of inlet spray nozzles to the existing cyclone. The capital cost for this method is based on 50 percent solids being fed to the existing Venturi. Any concentration less than 50 percent would require additional multiple effect evaporator capacity to increase the feed liquor concentration.

The existing tubular air heater will be replaced with a steam coil air heater. This replacement cost, as well as the cost for piping and duct changes, is included in



ADDED EQUIPMENT SHOWN WITH HEAVY LINES

FIG. 5 - 30 TYPICAL ARRANGEMENT FOR CONVERTING EXISTING VENTURI TO DIRECT CONTACT EVAPORATOR AND ADDING PRECIPITATOR - RECOVERY BOILER

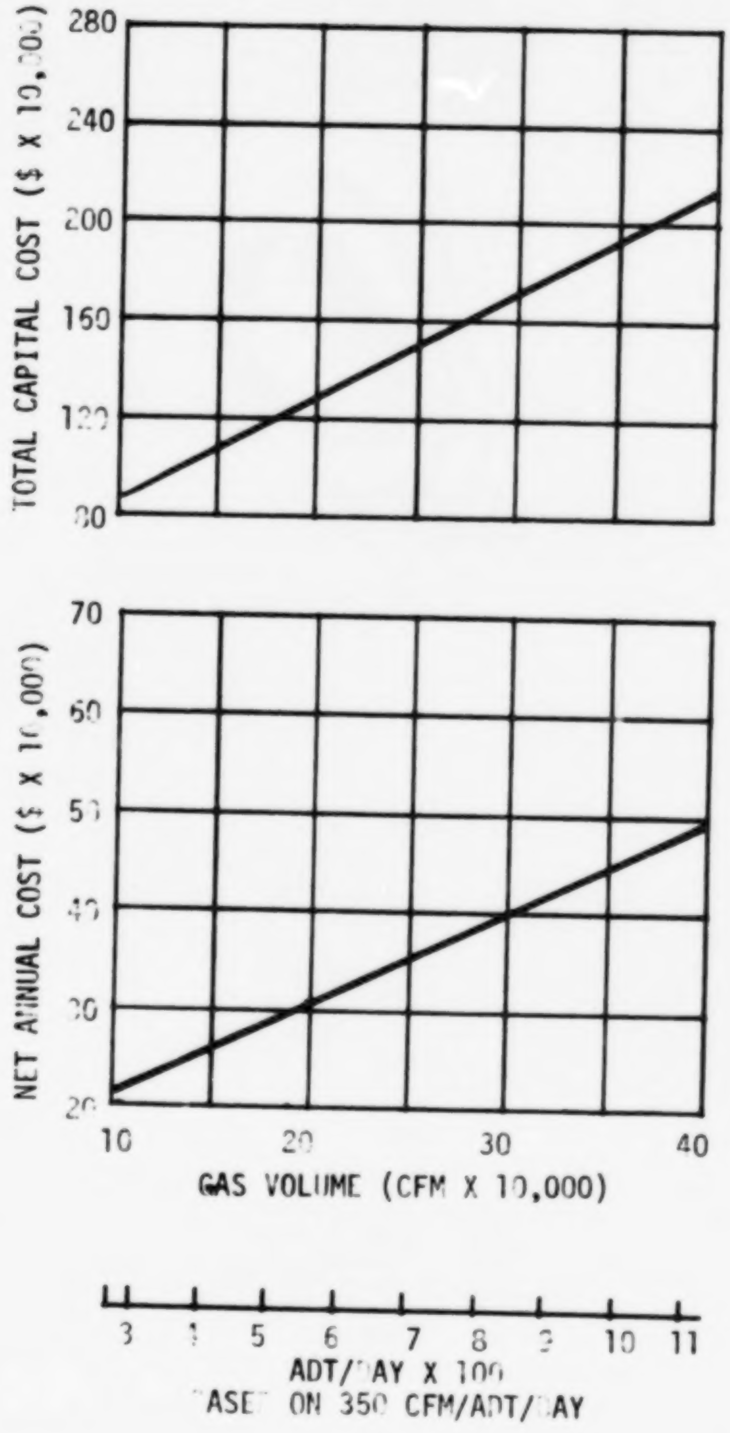


FIG. 5 - 31 CONTROL METHOD COSTS FOR CONVERTING EXISTING VENTURI TO CYCLONE EVAPORATOR AND ADD 99.9% EFFICIENCY PRECIPITATOR - RECOVERY BOILER

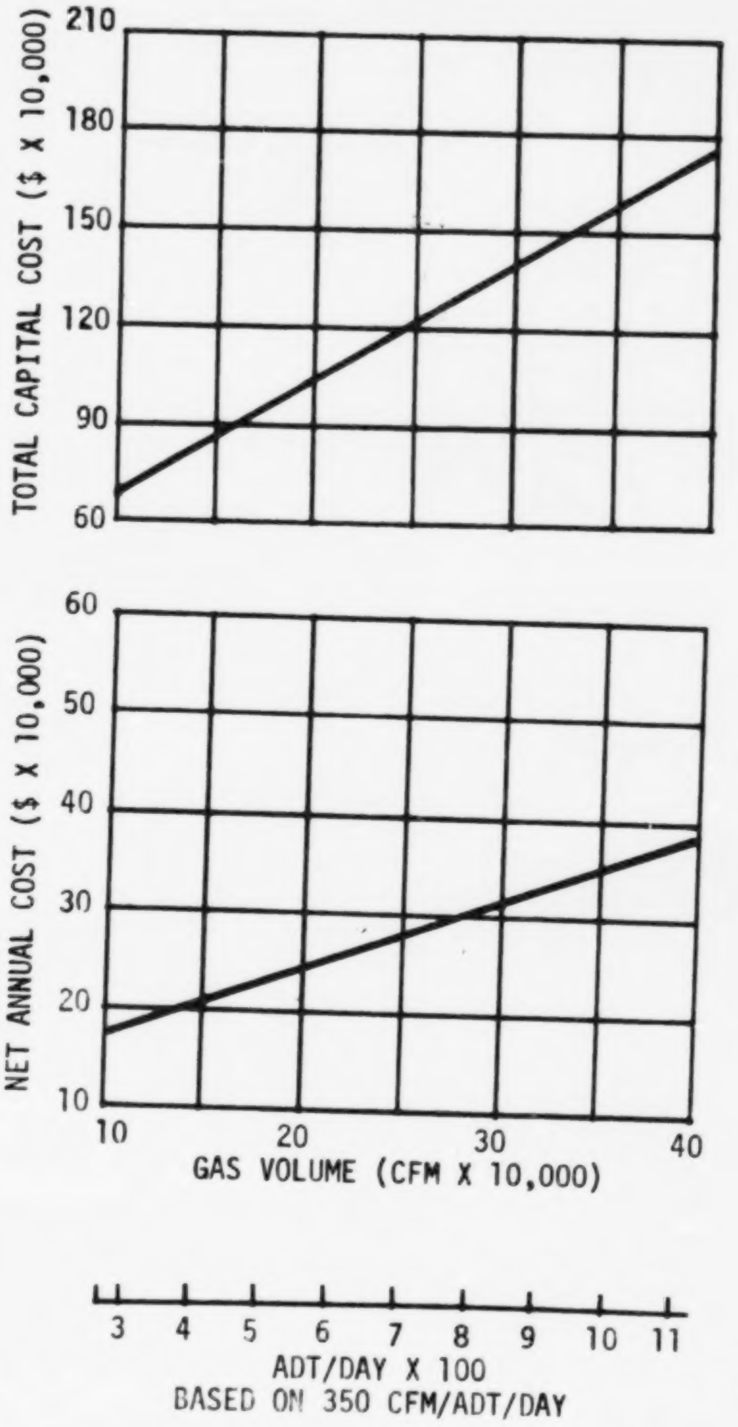


FIG. 5 - 32 CONTROL METHOD COSTS FOR CONVERTING EXISTING VENTURI TO CYCLONE EVAPORATOR AND ADD 99.0% EFFICIENCY PRECIPITATOR-RECOVERY BOILER

the capital cost estimate. The cost for additional steam requirement for the steam coil air heater is reflected in the annual net cost for this method.

A new steel stack is included in the capital cost.

Second Stage Venturi. This method is based on a second stage Venturi scrubber being installed subsequent to an existing Venturi as shown in Figure 5-33. The existing Venturi will be converted to a low energy Venturi and used as an evaporator. To complete this arrangement, a new I.D. fan with a greater statis head, a new recirculation pump, a stack, and revised 304 stainless steel ductwork from the existing Venturi to the Venturi scrubber and from the Venturi to the stack will be required. The second stage Venturi will utilize fresh water as the scrubbing medium. The scrubber effluent will be pumped to the first stage Venturi as shown in the flow schematic for this control method.

Costs as shown in Figure 5-34 are based on the above equipment and the instruments, motor controls, piping and power wiring required for proper operation. The net annual cost shown reflects a large credit due to chemical recovery. This credit is based on the increase of an operating collection efficiency from 80 to 97 percent AOE. For an actual case with a more efficient existing system and an existing higher system pressure drop, the credit for chemical recovery would decrease, a power cost savings may be realized, and the existing I.D. fan may be adequate, thus reducing the capital cost.

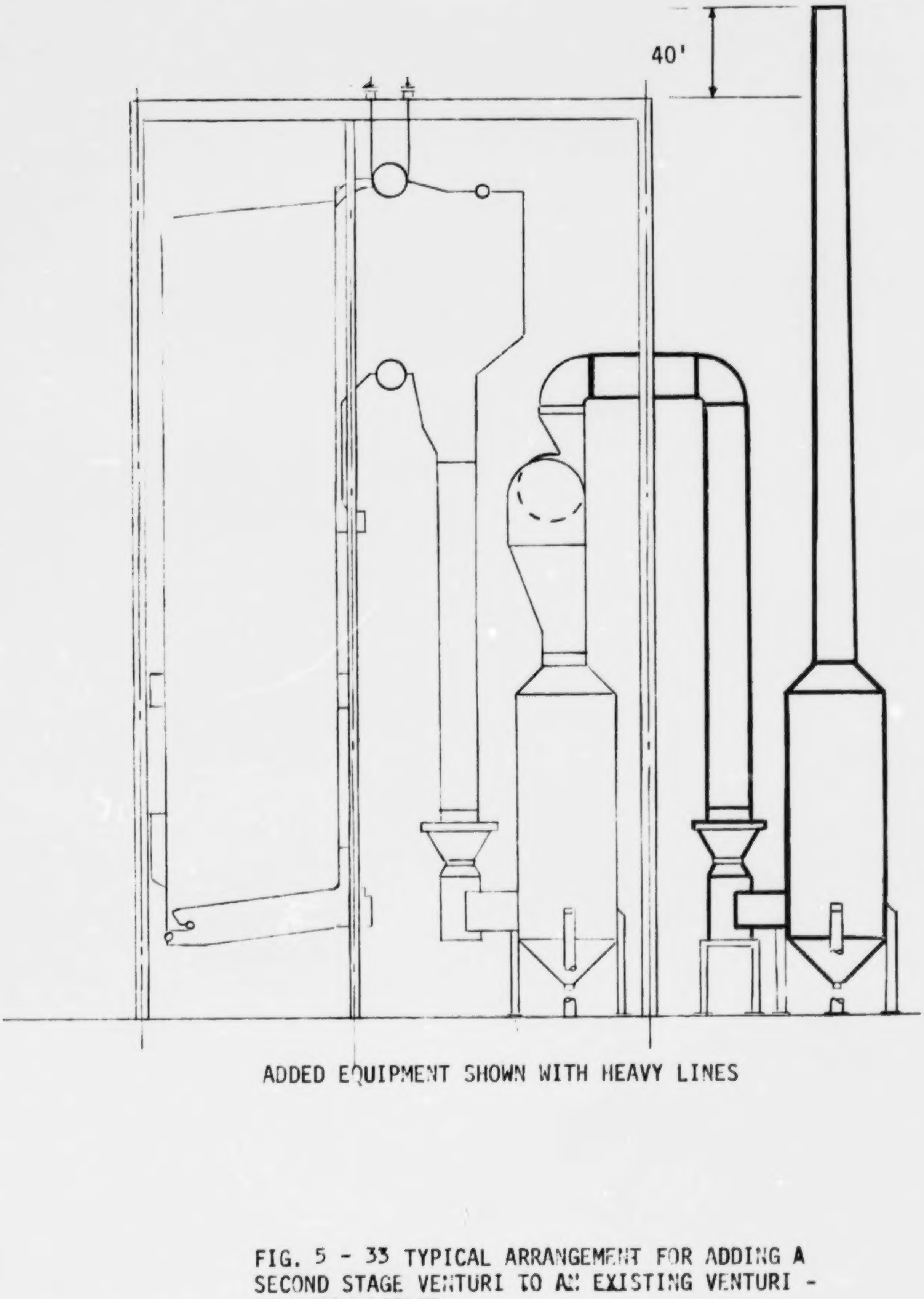
Effectiveness

Particulate Removal. Efficiencies for removal of particulate are as follows (based on 2 grains per ACF):

Control Method	Guaranteed Efficiency	Annual Operating Efficiency
Precipitator	99.9	99.5
Precipitator	99.0	98.5
Two Stage Venturi	98.0*	97.0

Curves for particulate cost effectiveness are shown in Figures 5-35 and 5-36.

*The two stage Venturi has a guaranteed efficiency of 99.0 percent based on the 4 grain per ACF particulate loading leaving the recovery economizer. This 99.0 percent corresponds to a guaranteed efficiency of 98.0 percent when adjusted for a 2 grain per ACF particulate loading, leaving a conventional DCE.



ADDED EQUIPMENT SHOWN WITH HEAVY LINES

FIG. 5 - 33 TYPICAL ARRANGEMENT FOR ADDING A SECOND STAGE VENTURI TO AN EXISTING VENTURI - RECOVERY BOILER

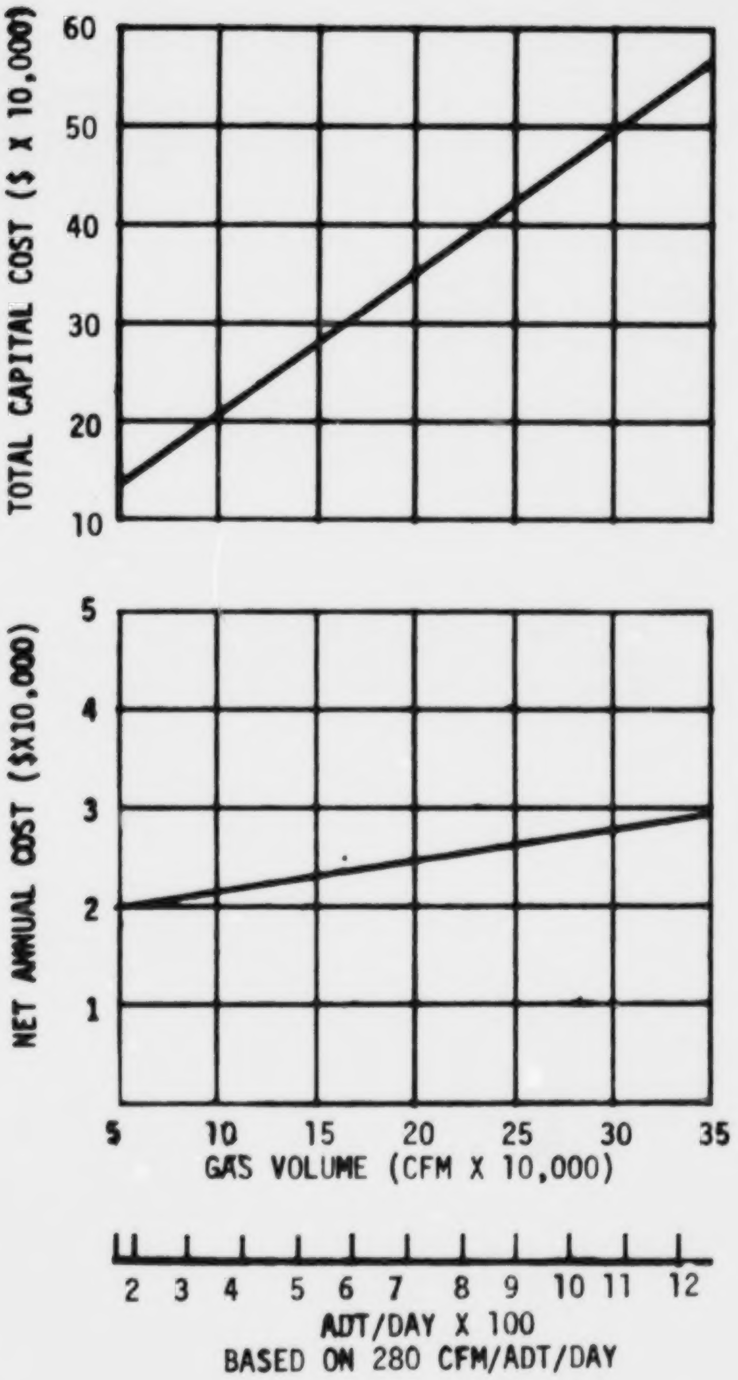
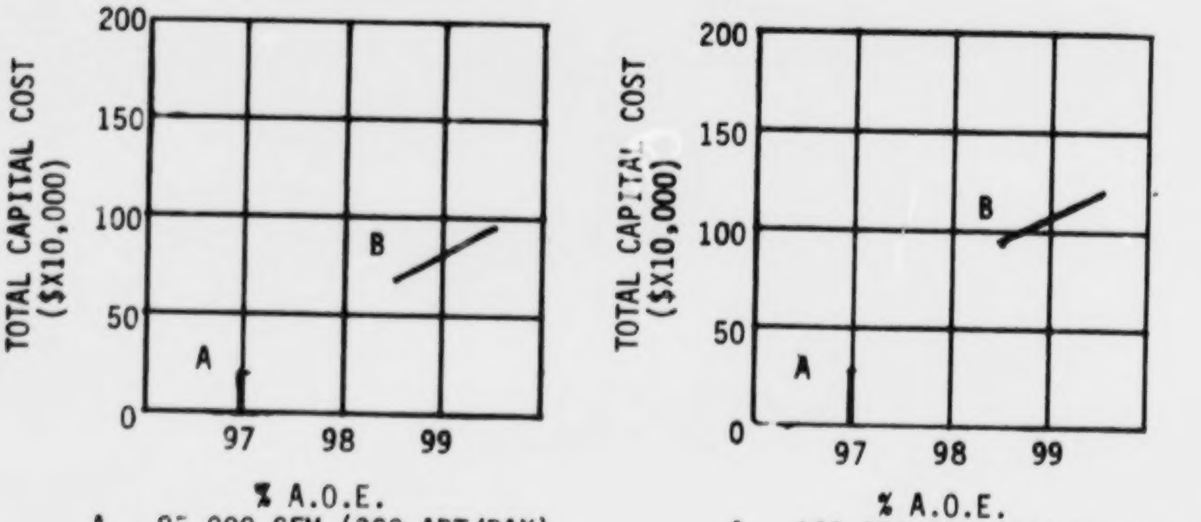
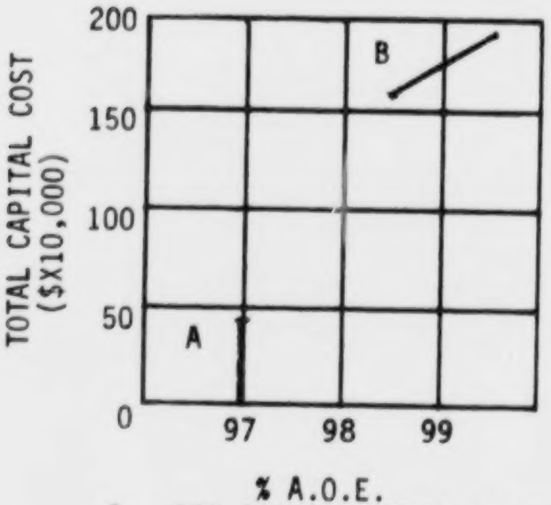


FIG.5-34 CONTROL METHOD COSTS FOR ADDING A SECOND STAGE VENTURI SCRUBBER TO AN EXISTING VENTURI - RECOVERY BOILER (97% A.O.E.)



A. 85,000 CFM,(300 ADT/DAY)
B. 105,000 CFM,(300 ADT/DAY)

A. 140,000 CFM,(500 ADT/DAY)
B. 175,000 CFM,(500 ADT/DAY)



A. 280,000 CFM,(1000 ADT/DAY)
B. 350,000 CFM,(1000 ADT/DAY)

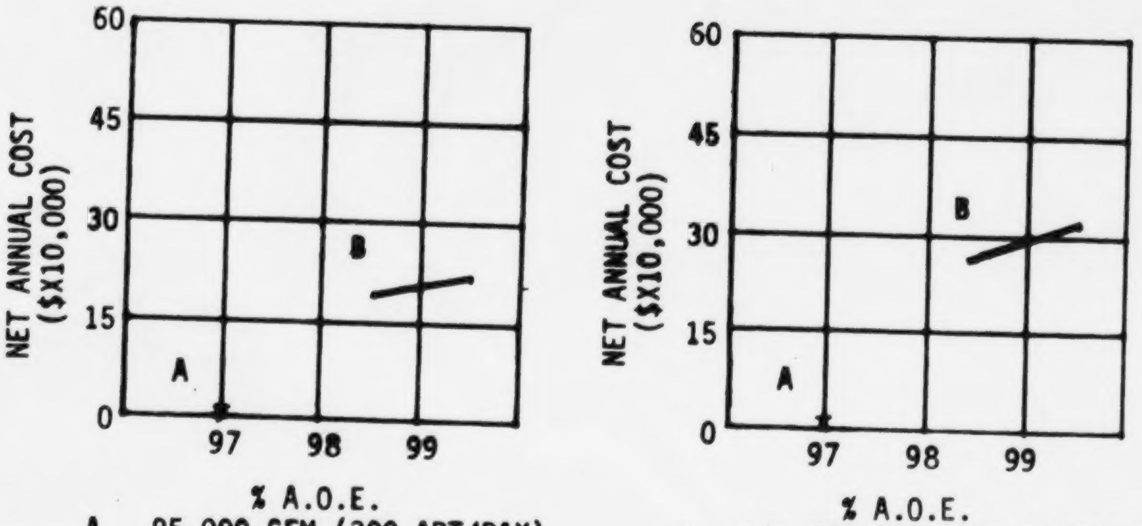
A. ADDITION OF A SECOND STAGE VENTURI TO AN EXISTING VENTURI

NOTE - ADT/DAY BASED ON 280 CFM/ADT/DAY

B. CONVERT EXISTING VENTURI EVAPORATOR TO CYCLONE EVAPORATOR AND ADD AN ELECTROSTATIC PRECIPITATOR

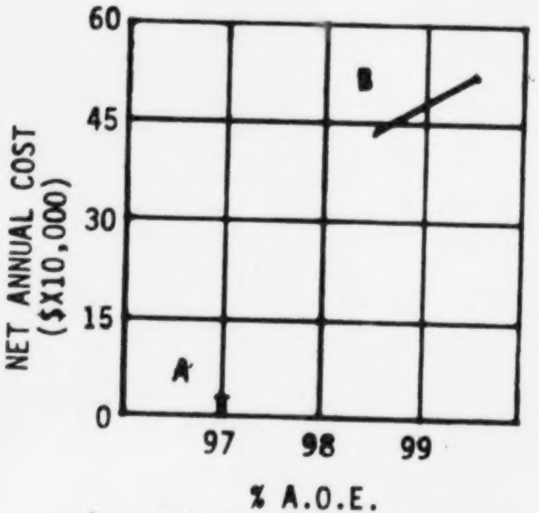
NOTE - ADT/DAY BASED ON 350 CFM/ADT/DAY

FIG.5-35 PARTICULATE COST EFFECTIVENESS (TOTAL CAPITAL COST COMPARISON) FOR INCREASING THE EFFICIENCY OF AN EXISTING 80% A.O.E. VENTURI EVAPORATOR SYSTEM - RECOVERY BOILER



A. 85,000 CFM,(300 ADT/DAY)
B. 105,000 CFM,(300 ADT/DAY)

A. 140,000 CFM,(500 ADT/DAY)
B. 175,000 CFM,(500 ADT/DAY)



A. 280,000 CFM,(1000 ADT/DAY)
B. 350,000 CFM,(1000 ADT/DAY)

A. ADDITION OF A SECOND STAGE VENTURI TO AN EXISTING VENTURI

NOTE - ADT/DAY BASED ON 280 CFM/ADT/DAY

B. CONVERT EXISTING VENTURI EVAPORATOR TO CYCLONE EVAPORATOR AND ADD AN ELECTROSTATIC PRECIPITATOR

NOTE - ADT/DAY BASED ON 350 CFM/ADT/DAY

FIG.5-36 PARTICULATE COST EFFECTIVENESS (NET ANNUAL COST COMPARISON) FOR INCREASING THE EFFICIENCY OF AN EXISTING 80% A.O.E. VENTURI EVAPORATOR SYSTEM - RECOVERY BOILER

Reduced Sulfur and Sulfur Dioxide Removal. If the recovery unit is operated at complete combustion conditions, the removal of the Venturi would not significantly affect the emissions of reduced sulfur compounds.

However, the removal of the Venturi would probably increase the SO_2 emission somewhat. Limited in-house test data indicate that the SO_2 emission might increase from an approximate range of 1/2 - 10 lbs/ADT to an approximate range of 1 - 15 lbs/ADT, when the Venturi is removed.

Operation. In comparing operation, the second stage Venturi method would result in more potential corrosion and a lower plume rise.

Summary. The second stage Venturi method has the lowest cost; however, the efficiency is limited to 97.0 AOE. If a higher efficiency is required, then the 99.5 AOE precipitator method should be considered.

5.3.6.1.5 Recovery System, Case 5

Application

This case is based on the installation of a black liquor oxidation system to reduce hydrogen sulfide emissions from an existing recovery system. Both weak and concentrated black liquor oxidation systems are considered. As discussed in Section 5.2.7, the selection of an oxidation system for a specific installation will depend to a large extent on the characteristics of the liquor. Flow diagrams for the systems considered are shown in Figures 5-37 and 5-38.

Costs

The total capital costs of weak and also concentrated liquor oxidation systems appear on the cost curves, Figures 5-39 and 5-40. Air requirements for the weak liquor system are shown in Figure 5-41. These total capital costs consist of the total direct capital costs, plus indirect capital costs. The indirect capital costs are computed as 30 percent of the direct cost. Included in the total direct capital costs are: Equipment and equipment erection (which includes instrumentation and motors), equipment foundations and buildings, process and instrument piping, power wiring and lighting.

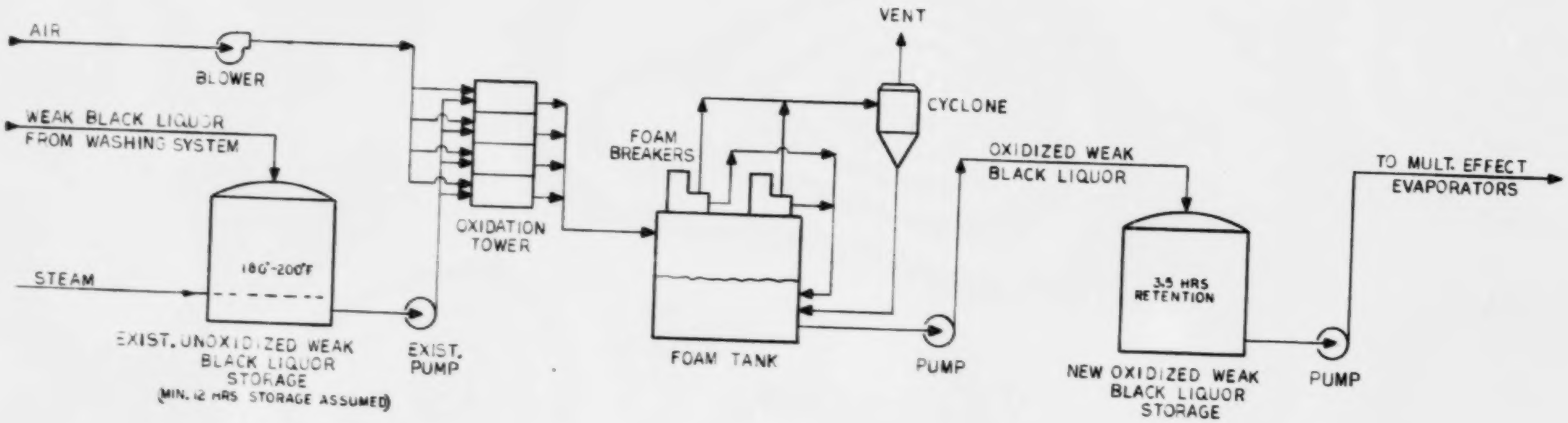

AIR
BLOWER
WEAK BLACK LIQUOR
FROM WASHING SYSTEM
STEAM
180°-200°F
EXIST. UNOXIDIZED WEAK
BLACK LIQUOR
STORAGE
(MIN. 12 HRS STORAGE ASSUMED)
EXIST.
PUMP
OXIDATION
TOWER
FOAM
BREAKERS
FOAM TANK
VENT
CYCLONE
PUMP
OXIDIZED WEAK
BLACK LIQUOR
3.5 HRS
RETENTION
NEW OXIDIZED WEAK
BLACK LIQUOR
STORAGE
PUMP
TO MULT. EFFECT
EVAPORATORS

FIG. 5-37

WEAK LIQUOR OXIDATION SYSTEM

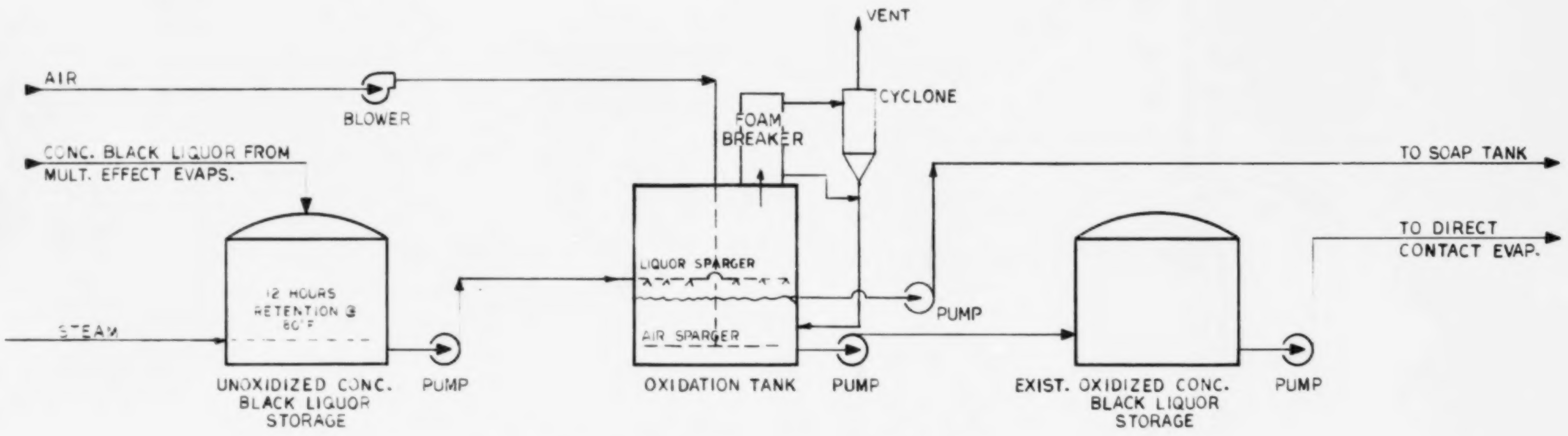
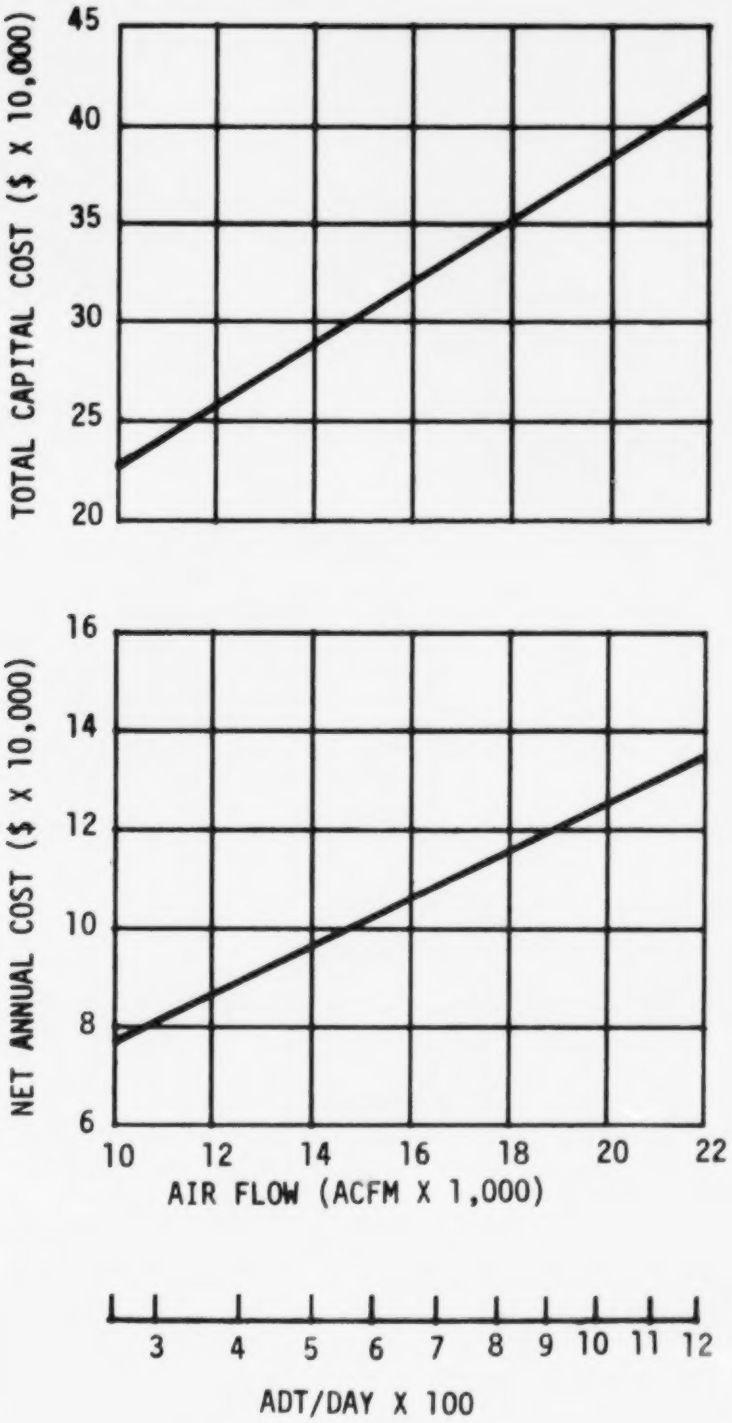

AIR
BLOWER
CONC. BLACK LIQUOR FROM
MULT. EFFECT EVAPS.
STEAM
12 HOURS
RETENTION @
80°F
UNOXIDIZED CONC.
BLACK LIQUOR
STORAGE
PUMP
VENT
CYCLONE
FOAM
BREAKER
LIQUOR SPARGER
AIR SPARGER
OXIDATION TANK
PUMP
PUMP
TO SOAP TANK
EXIST. OXIDIZED CONC.
BLACK LIQUOR
STORAGE
PUMP
TO DIRECT
CONTACT EVAP.

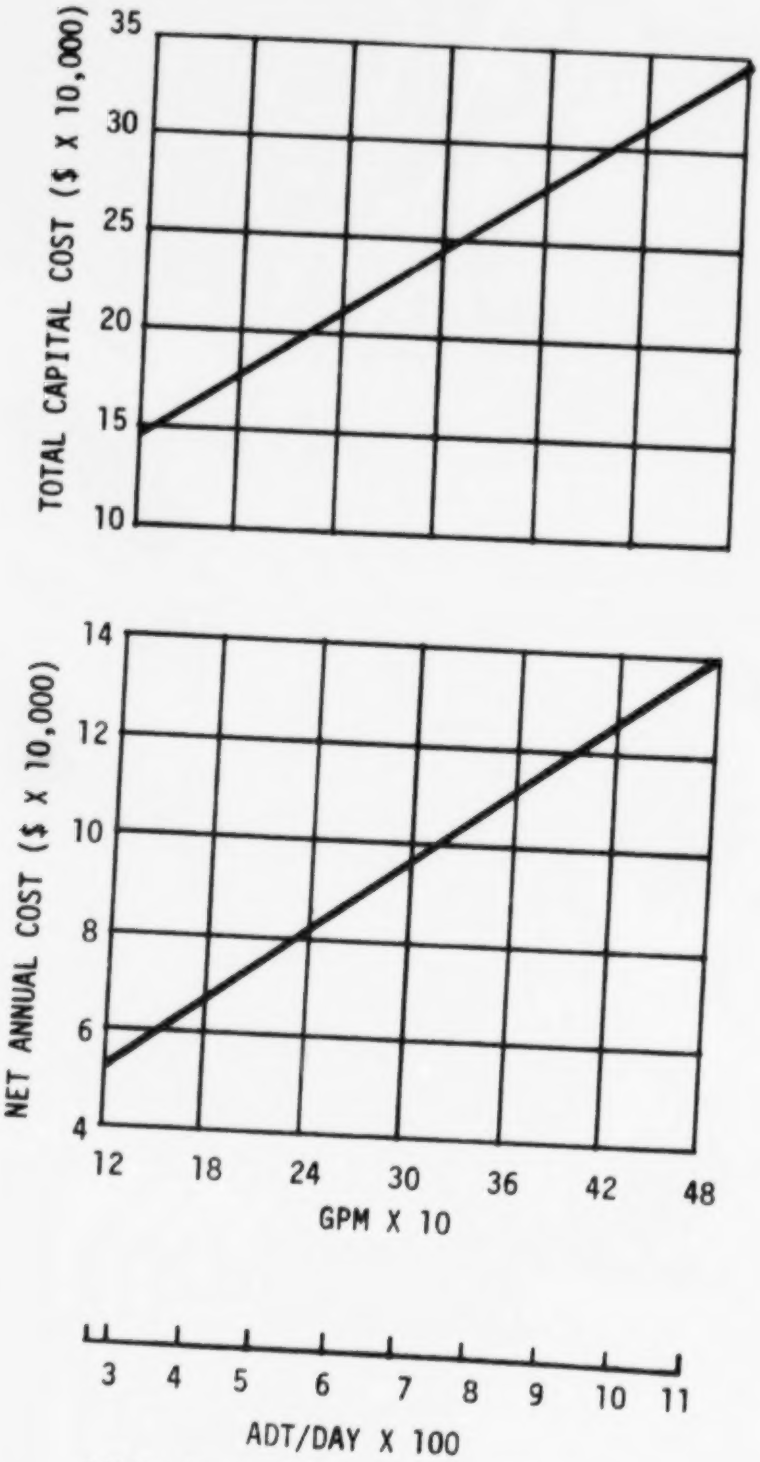
FIG. 5-38

CONCENTRATED LIQUOR OXIDATION SYSTEM



BASED ON 3,000 LBS. B. L. SOLIDS/ADT, 15% B. L. SOLIDS, 10 GRAMS Na_2S/LITER, SPECIFIC GRAVITY OF 1.1

FIG. 5 - 39 CONTROL METHOD COST FOR WEAK BLACK LIQUOR OXIDATION



BASED ON 3,000 LBS. B.L. SOLIDS/ADT, 50% B.L. SOLIDS, SPECIFIC GRAVITY OF 1.25

FIG. 5 - 40 CONTROL METHOD COSTS FOR CONCENTRATED BLACK LIQUOR OXIDATION

WEAK BLACK LIQUOR OXIDATION
AIR REQUIREMENTS

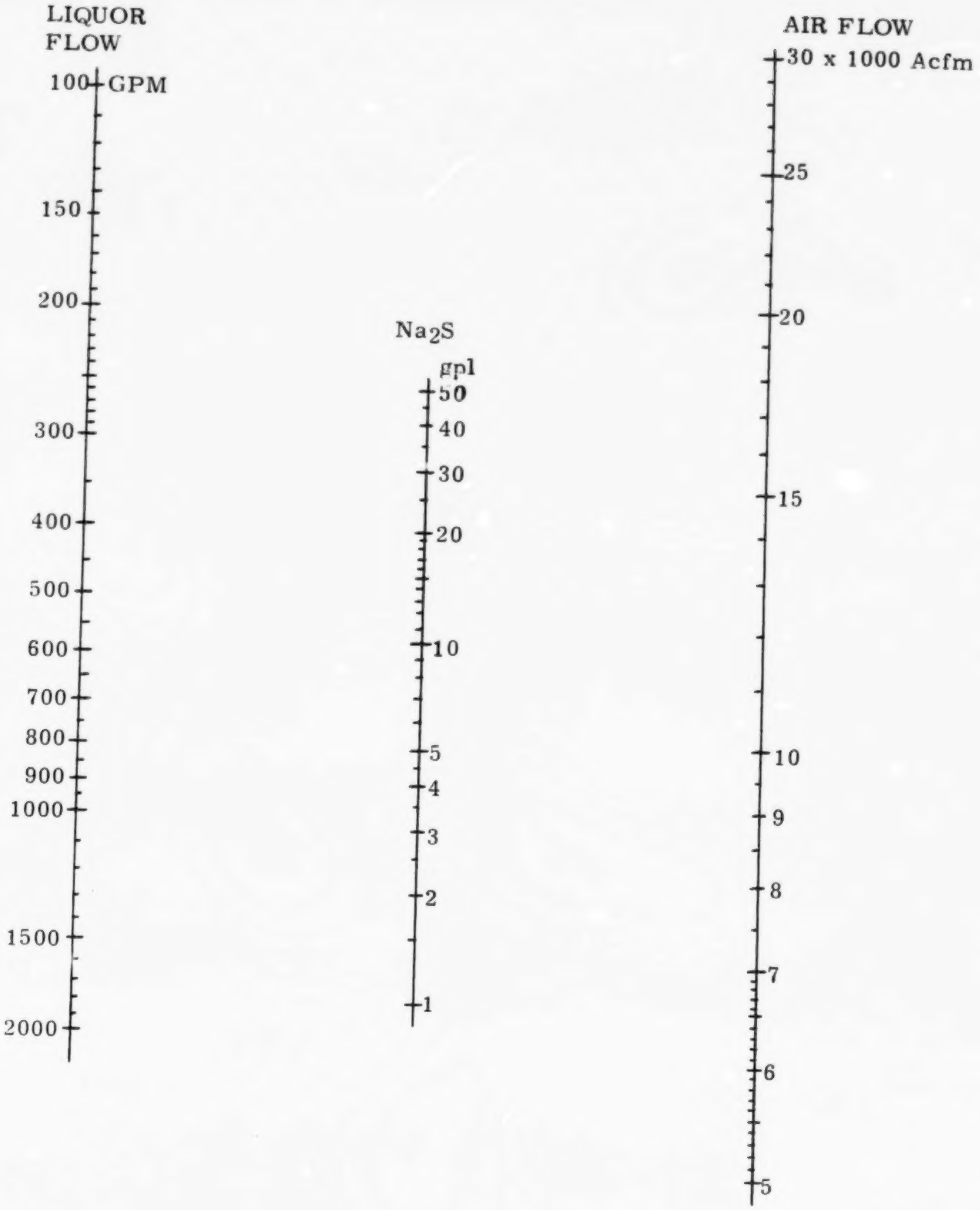


FIG. 5 - 41

Effectiveness

Particulate. There are usually no particulate emissions from the oxidations systems. Oxidation of liquor has no effect on particulate emissions from the recovery system.

Reduced Sulfur and Sulfur Dioxide Removal. The basic purpose of an oxidation system is to reduce the malodorous emission from the direct contact evaporator. This malodorous emission from the direct contact evaporator primarily occurs in the form of reduced sulfur gases (H_2S, RSH, RSR, RSSR). As stated previously, when the Na_2S is oxidized to sodium thiosulfate ($Na_2S_2O_3$), the above reaction does not occur, since sodium thiosulfate is a relatively stable compound.

Of course, the oxidation of the liquor will not reduce the emission of odorous compounds from the recovery furnace. Odorous furnace emissions are a function of recovery furnace operation, such as overload, air distribution, and spray droplet size. Oxidation will only reduce emissions from the direct contact evaporator.

Experience at several mills has proven that any oxidation system must operate at the highest possible efficiency. The term efficiency is rather nebulous since the measure of Na_2S in the liquor is the true indicator of oxidation performance. Therefore, any oxidation system should be designed to provide liquor at a Na_2S content of 0.10 grams per liter or less to the direct contact evaporator.

At this level, the H_2S level in the flue gases should average 5 - 20 ppm by volume, if the furnace is operated properly. Upsets in furnace operation may extend this range as high as 100 ppm.

Concentrations of other reduced sulfur compounds are even more difficult to define than H_2S.

For high efficiency oxidation systems, the reduced sulfur compounds--RSR, RSH, RSSR--are present in very low concentrations. Most test results on properly operated furnaces report a range from 0 to 3 ppm leaving the direct contact evaporator.

Since H_2S has been the most predominant odorous compound leaving the direct contact evaporator, the other reduced sulfur compounds have been viewed in the past as secondary in importance. While this is true, the resolution of H_2S emissions will require increasing emphasis to be directed to the remaining reduced sulfur compounds. Perhaps, the emissions of these compounds will become more fully understood in the future.

The role of sulfur dioxide emission in relation to black liquor oxidation has not been clearly defined. However, actual experience indicates that SO_2 emission after oxidation (based upon proper furnace operations) will not exceed the SO_2 loss before oxidation. Mills which operate at a low sodium sulfide residual to the direct contact evaporator have reported SO_2 levels in the order of 100 ppm by volume.

Operation. The major emissions from the oxidation system stack are dimethyl sulfide and dimethyl disulfide. The dimethyl sulfide, dissolved in the liquor, escapes to the atmosphere. On the other hand, the dimethyl disulfide is created by the oxidation of methyl mercaptan by oxygen in the presence of alkali, the dimethyl disulfide then being released to the atmosphere.

Maintenance costs for the two oxidation systems are not clearly defined due to many design variations and the unavailability of adequate maintenance cost records. Since the capital costs of the two systems are relatively equal, the maintenance costs are also assumed to be approximately equal.

For properly designed and applied systems, the reliability of weak or concentrated oxidation systems are approximately equal.

Summary. The capital costs and net annual costs for this method do not include any allowance for equipment to treat the exhaust gas from the oxidation systems. Further, the net annual cost does not include any allowance for chemical recovery, since this savings (if any) will vary from mill to mill and must be individually calculated. Provisions should be considered for treatment of this exhaust gas, where minimum emission is essential.

In current technology, oxidation systems must compete with new designs which eliminate direct contact between the black liquor and the flue gases to determine the overall recovery system most suitable for a particular mill. A comparison of these new systems applied to existing mills is discussed in Section 6.2.1.1. This case is based only on an appraisal of weak liquor oxidation versus concentrated liquor oxidation.

Cost curves have been prepared showing net annual cost versus tons per day production for weak and concentrated liquor oxidation systems in Figure 5-42.

As shown in the curves, the net annual costs of the two systems are very close, depending upon the sodium sulfide concentration. For liquor oxidation systems designed for 10 grams/liter of Na_2S, a concentrated system appears to be the best choice for sizes below 800 ADT/Day. Considering the fact that a concentrated system eliminates the sodium sulfide reversion reaction which takes place with a weak liquor system, a concentrated system should probably be selected even for tonnages higher than 800 ADT/Day.

In conclusion, the selection of the oxidation system must be analyzed for each mill; however, considering all factors, a concentrated system appears to be the best choice.

5.3.6.1.6 Recovery System, Case 6

Application

For older recovery systems (more than ten years old), odorous emissions from the recovery furnace may be significant, even though the most modern current technology (proper operation and black liquor oxidation) is employed. It is conceivable that for some older recovery systems, the odorous emissions may never be reduced to levels which are currently being considered by regulatory authorities (this refers to H_2S levels in the order of 17 ppm). This may be attributed to the older furnace configurations and the older systems which introduce air into the recovery furnace. Therefore, these older recovery systems may have to be replaced with one of the recovery systems listed below. Accordingly, approximate capital costs have been prepared for a range of sizes of the following recovery systems:

1. New Conventional Recovery System (Figure 5-43)
2. New Conventional Recovery System with Concentrated Black Liquor Oxidation (Figure 5-43)

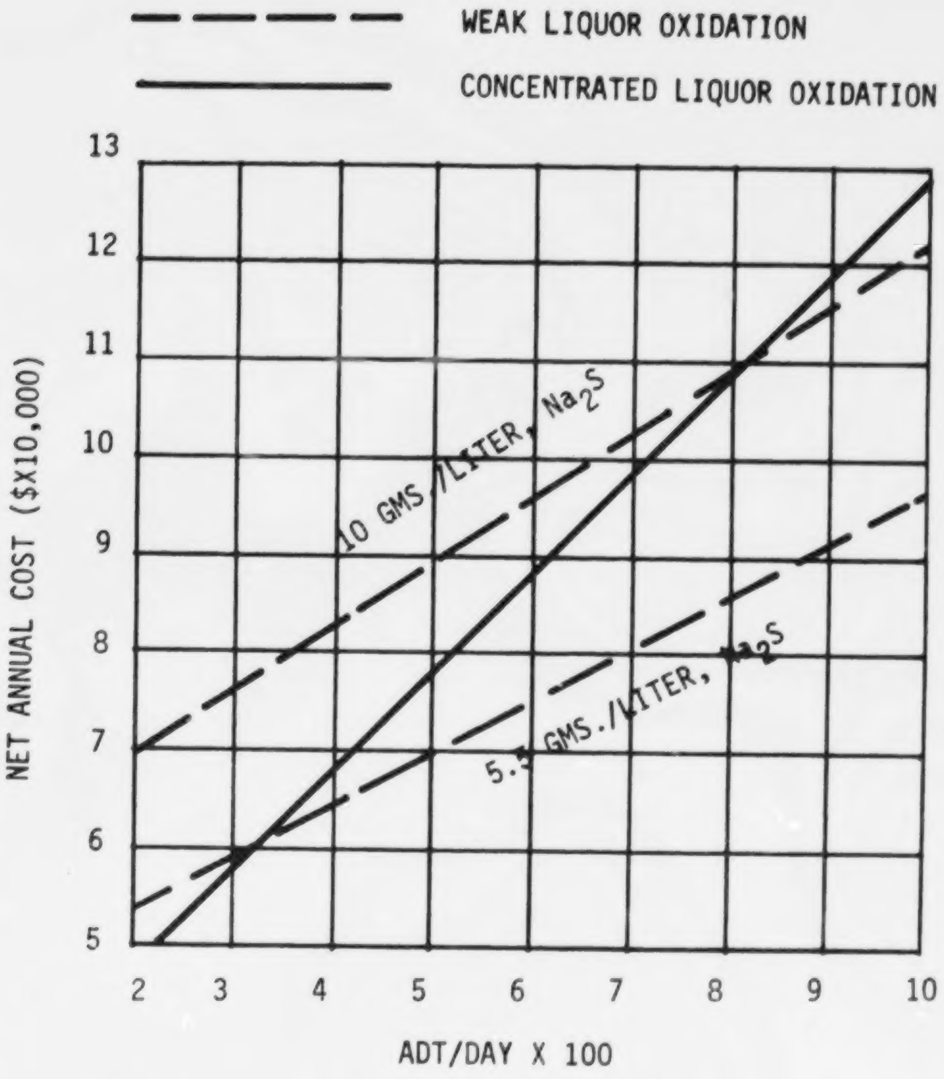


FIG.5 - 42 COMPARISON OF NET ANNUAL COSTS FOR BLACK LIQUOR OXIDATION SYSTEMS

3. New Recovery System Utilizing 62 Percent Solids Black Liquor from the Multiple Effect Evaporators (Figure 5-44)

4. New Air Contract Evaporator Recovery System (Figure 5-44)

Costs

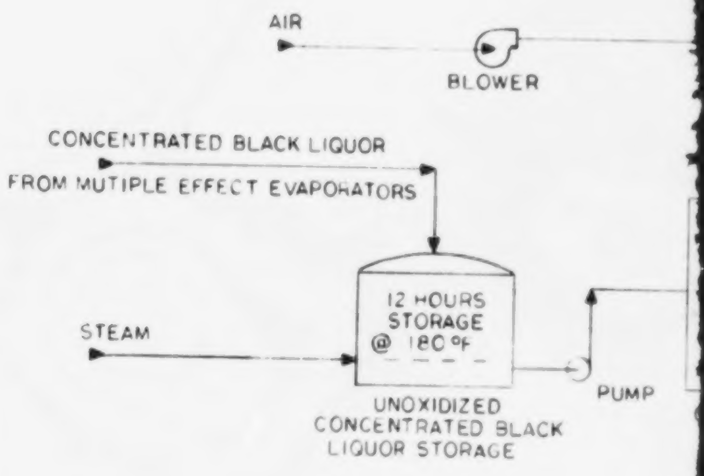
In estimating capital costs, the recovery system tonnages were based on 3,000 pounds of dry solids per ADT and 6,600 BTU's per pound of dry solids. Capital cost was estimated for two steam conditions of 600 psig-750°F and 1250 psig-900°F. Capital costs for 850 psig-825°F would be approximately midway between these two curves and may be interpolated.

The above capital costs are based on the U.S. average labor costs (U.S. Avg. = 100), and should be adjusted for the area of the country in regard to construction labor costs. This can be accomplished by adjusting 50 percent of the capital cost (labor and material comprise approximately 50 percent each of the capital cost), in relation to the Construction Labor Cost Index, Figure 5-45. Even after this adjustment, the capital cost may vary ± 10 percent depending upon geographical area (primarily due to weather considerations), company preferences, and differences in engineering design.

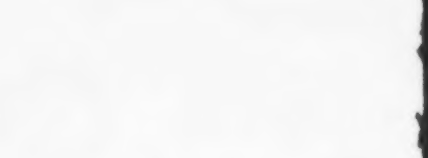
These capital costs assume that the new recovery system will replace an existing recovery system which is being operated at complete combustion conditions (approximately design capacity). Therefore, it is anticipated that only minor quantities of additional make-up water, cooling water, feedwater, compressed air, electricity, et cetera, will be required for this new system. If any of these utilities may be required in significant quantity, they must be added to the capital cost in this section. Further, these costs are based on an open area and do not include any allowance for site clearance. The cost of loss of production due to downtime for the installation of the new recovery system is not included.

Generally speaking, these capital costs may be described as follows:

1. A recovery system designed to produce steam at the turbine-generator operating conditions specified on the cost curves

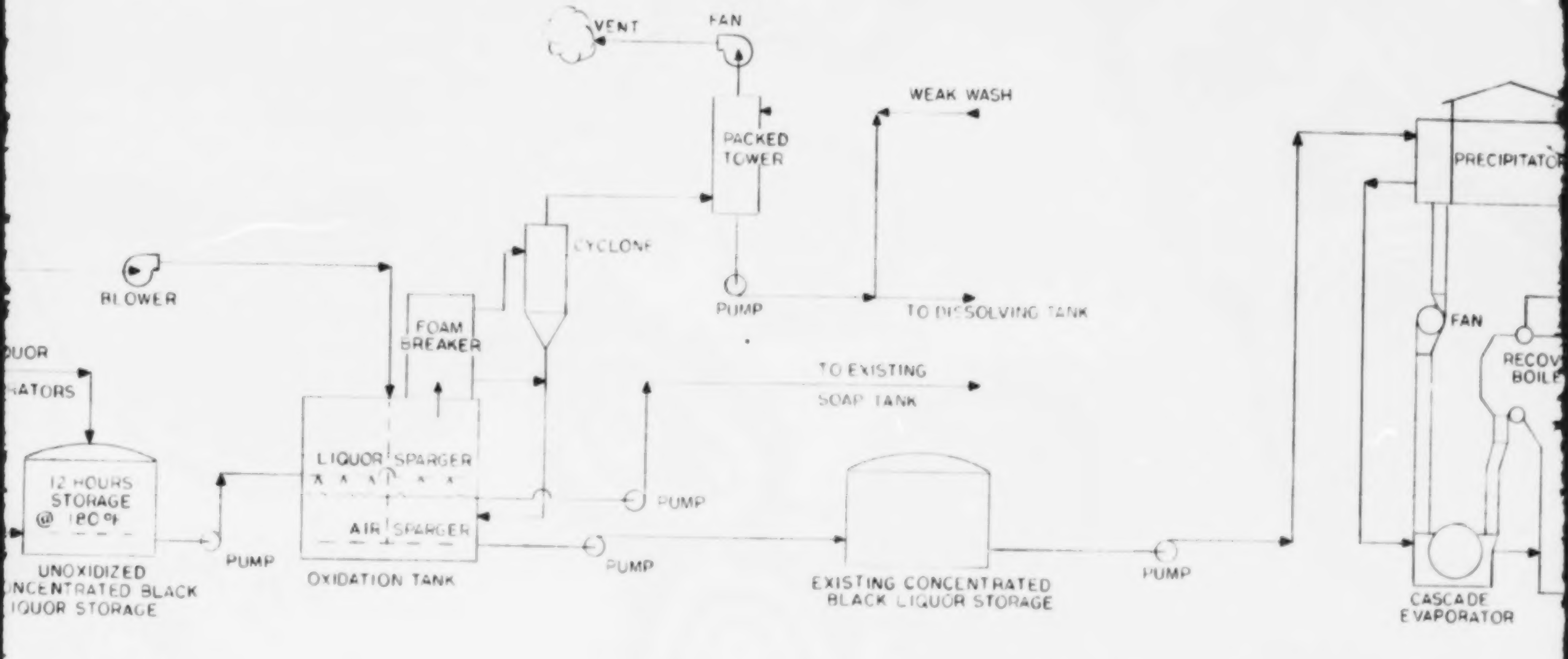

AIR
BLOWER
CONCENTRATED BLACK LIQUOR
FROM MUTIPLE EFFECT EVAPORATORS
12 HOURS STORAGE @ 180 °F
STEAM
PUMP
UNOXIDIZED CONCENTRATED BLACK LIQUOR STORAGE

CONVENTIONAL

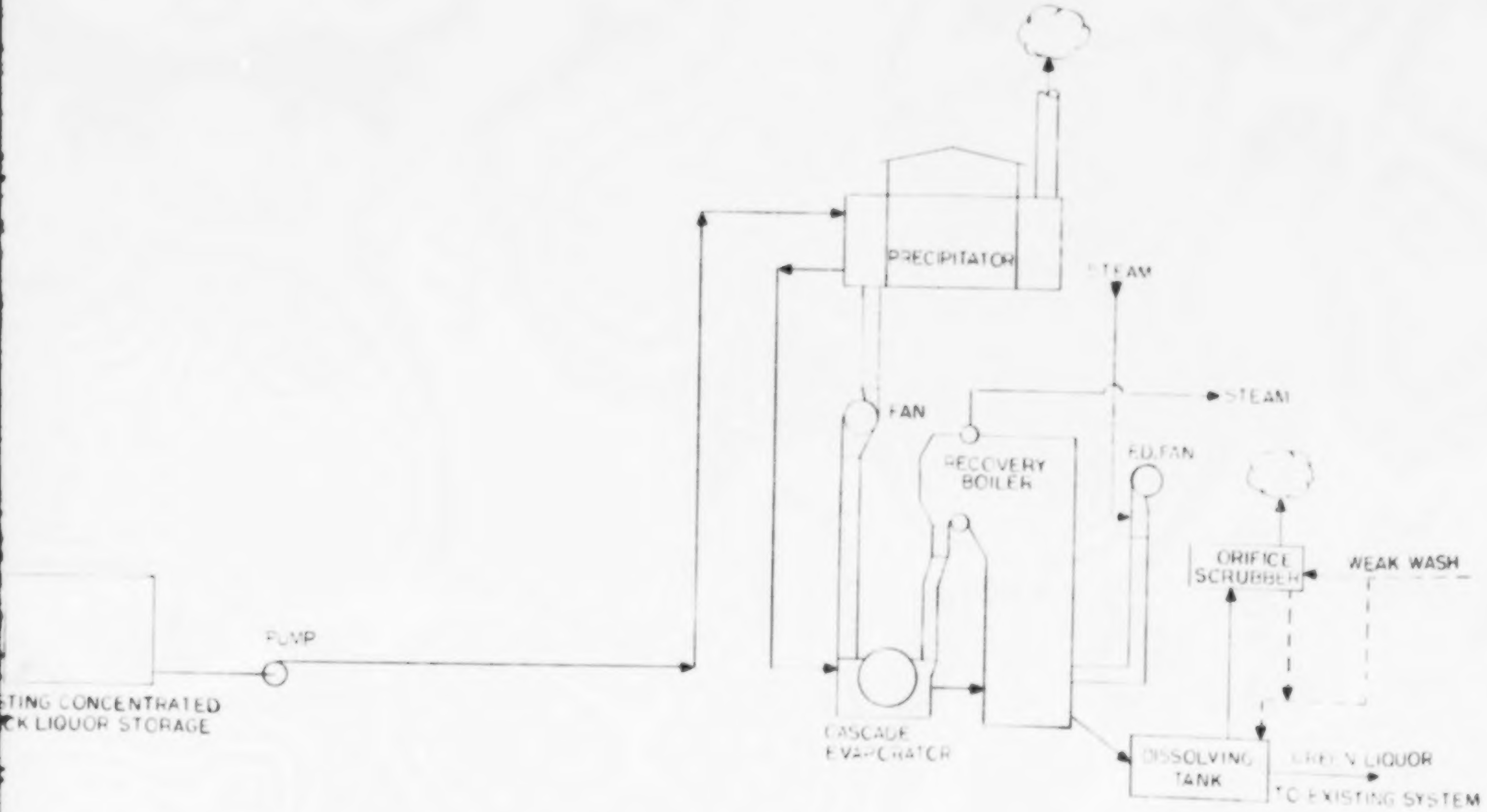

PUMP
EXISTING CONCENTRATED BLACK LIQUOR STORAGE

CONVENTI

A


VENT
FAN
WEAK WASH
PACKED TOWER
PRECIPITATOR
CYCLONE
BLOWER
PUMP
TO DISSOLVING TANK
FAN
FOAM BREAKER
UOR
RATORS
TO EXISTING
SOAP TANK
RECOV
BOILE
12 HOURS STORAGE @ 180 °F
LIQUOR SPARGER
AIR SPARGER
PUMP
PUMP
PUMP
PUMP
UNOXIDIZED
NCENTRATED BLACK
IQUOR STORAGE
OXIDATION TANK
EXISTING CONCENTRATED BLACK LIQUOR STORAGE
CASCADE EVAPORATOR

CONVENTIONAL KRAFT RECOVERY SYSTEM WITH CONCENTRATED BLACK LIQUOR OXIDATION


PRECIPITATOR
STEAM
STEAM
FAN
RECOVERY BOILER
I.D. FAN
ORIFICE SCRUBBER
WEAK WASH
PUMP
STING CONCENTRATED
CK LIQUOR STORAGE
CASCADE EVAPORATOR
DISSOLVING TANK
GREEN LIQUOR
TO EXISTING SYSTEM

CONVENTIONAL KRAFT RECOVERY SYSTEM

B

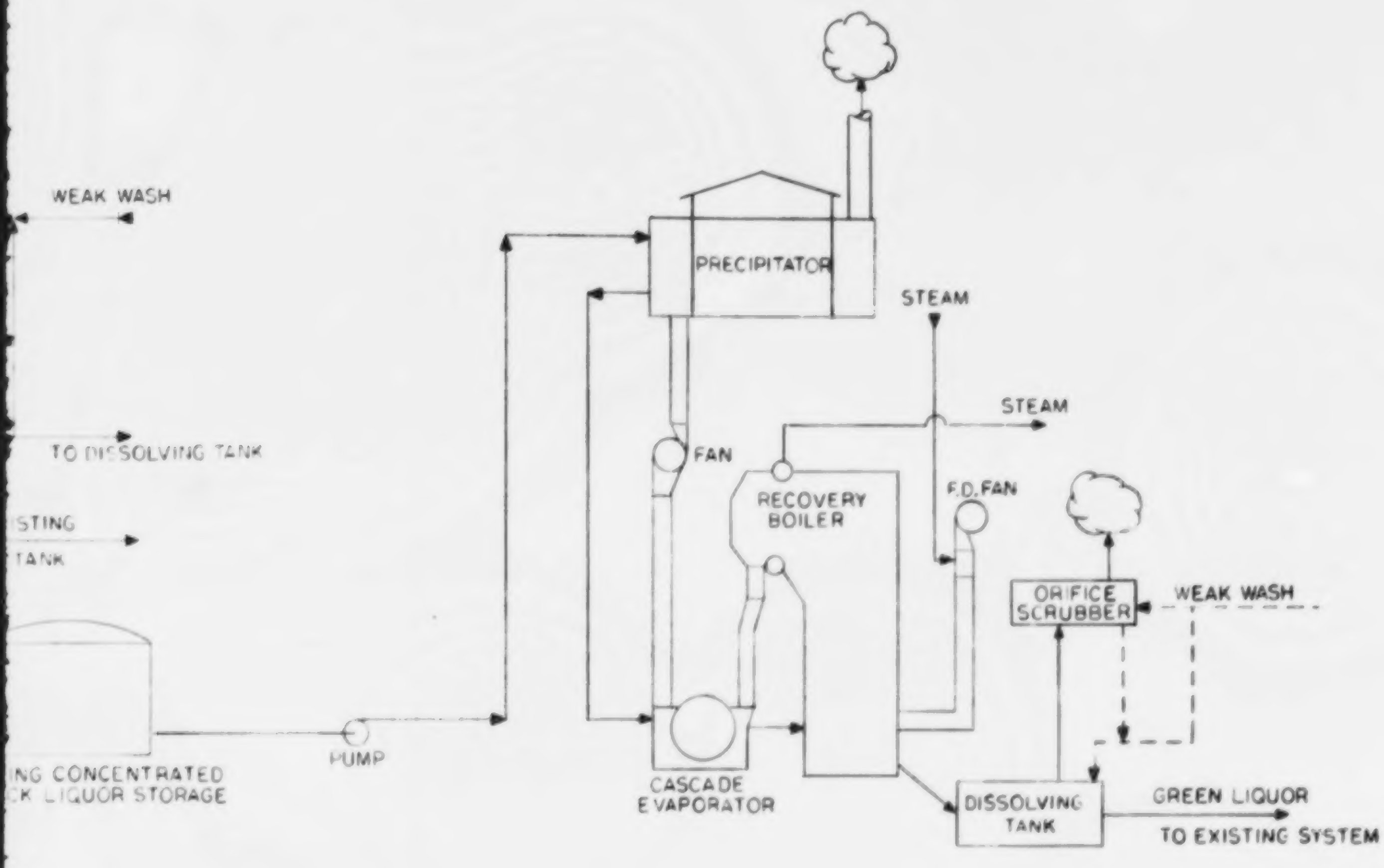
SCH
CONV. RECOVER
OXIDATION.

EMISS

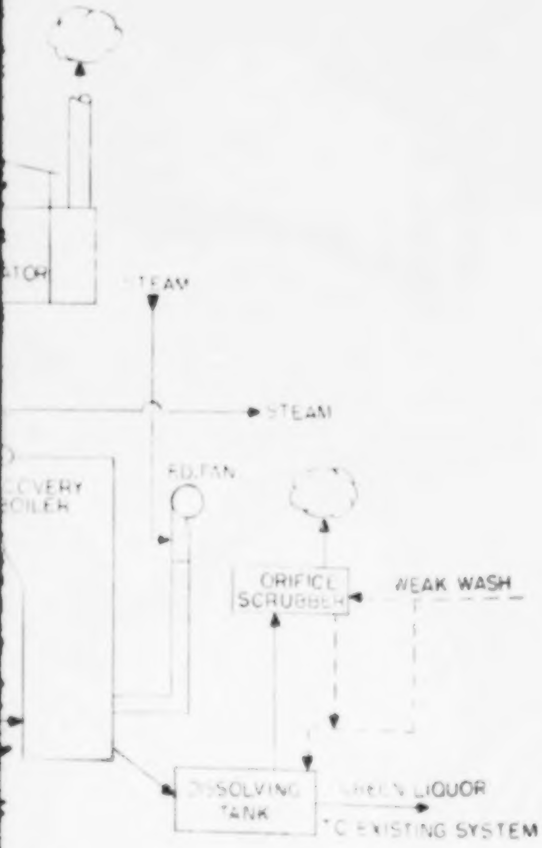
DEPA
CONSU
N

ENVIRONMENT
GAIN

1


WEAK WASH
TO DISSOLVING TANK
STING
TANK
ING CONCENTRATED
CK LIQUOR STORAGE
PUMP
PRECIPITATOR
STEAM
STEAM
FAN
RECOVERY
BOILER
F.D. FAN
ORIFICE
SCRUBBER
WEAK WASH
CASCADE
EVAPORATOR
DISSOLVING
TANK
GREEN LIQUOR
TO EXISTING SYSTEM

CONCENTRATED BLACK LIQUOR OXIDATION


ATOR
STEAM
STEAM
COVERY
OILER
F.D. FAN
ORIFICE
SCRUBBER
WEAK WASH
DISSOLVING
TANK
GREEN LIQUOR
TO EXISTING SYSTEM

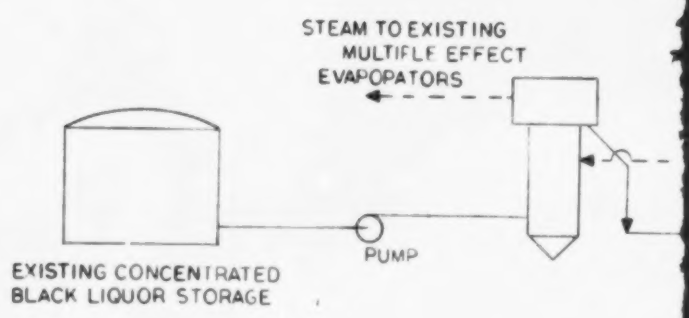
SCHEMATIC ARRANGEMENTS
CONV. RECOVERY AND CONV. RECOVERY WITH BLACK LIQUOR OXIDATION.

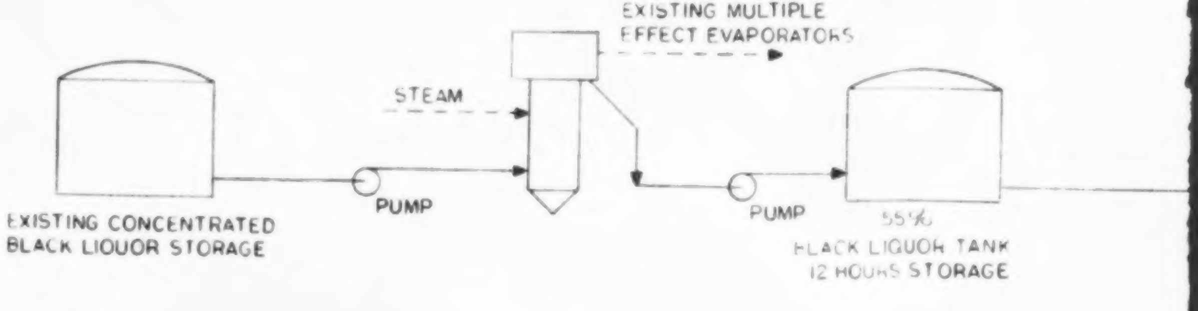
EXHIBIT NO.
FIG. 5-43
HJS

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

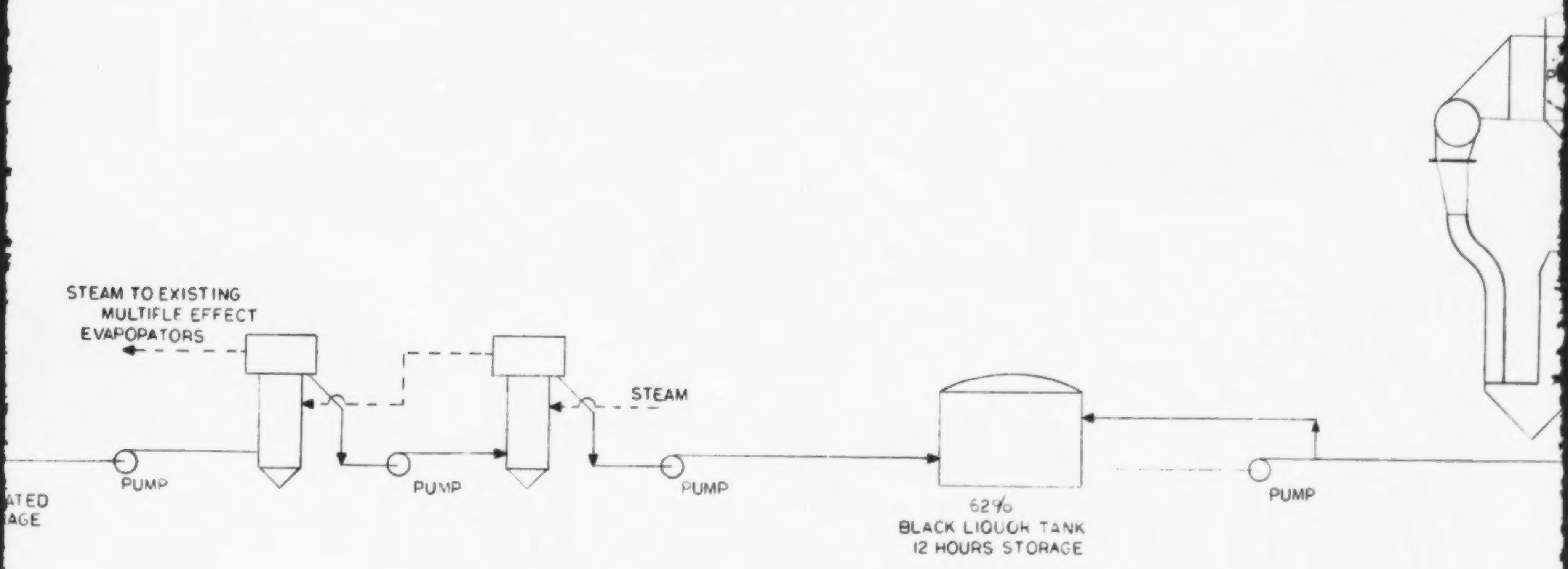
J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


STEAM TO EXISTING
MULTIFLE EFFECT
EVAPOPATORS
EXISTING CONCENTRATED
BLACK LIQUOR STORAGE
PUMP

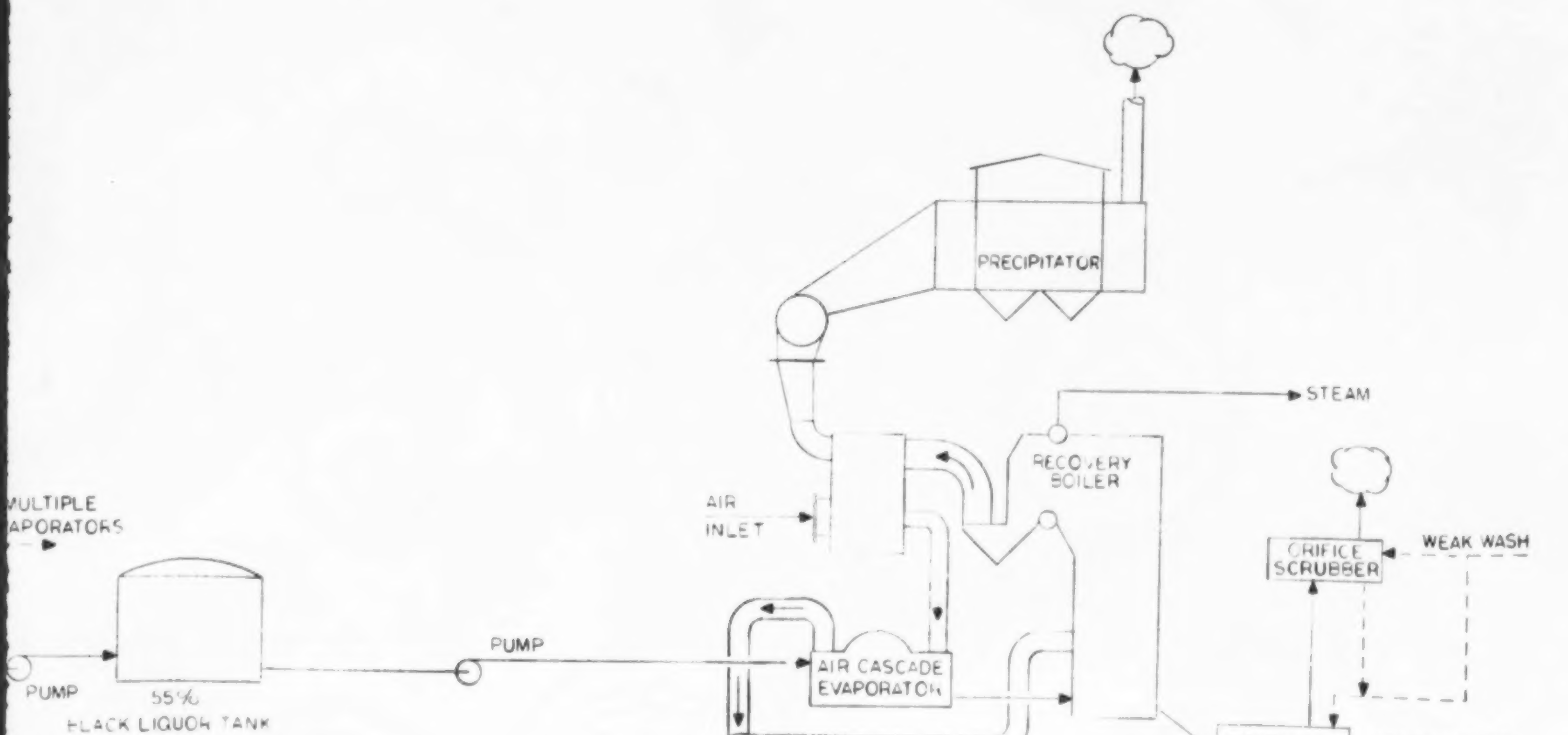

STEAM TO
EXISTING MULTIPLE
EFFECT EVAPORATORS
STEAM
EXISTING CONCENTRATED
BLACK LIQUOR STORAGE
PUMP
PUMP
55%
BLACK LIQUOR TANK
12 HOURS STORAGE

AIR CASCADE EVAPOR

A


STEAM TO EXISTING
MULTIFLE EFFECT
EVAPOPATORS
STEAM
ATED
AGE
PUMP
PUMP
PUMP
62%
BLACK LIQUOR TANK
12 HOURS STORAGE
PUMP

62% BLACK LIQUOR SOLIDS KRAFT RECOVERY SYSTEM


PRECIPITATOR
STEAM
RECOVERY
BOILER
AIR
INLET
MULTIPLE
APORATORS
ORIFICE
SCRUBBER
WEAK WASH
PUMP
55%
BLACK LIQUOR TANK
12 HOURS STORAGE
PUMP
AIR CASCADE
EVAPORATOR
FD
FAN
DISSOLVING
TANK
GREEN LIQUOR
TO EXISTING SYSTEM

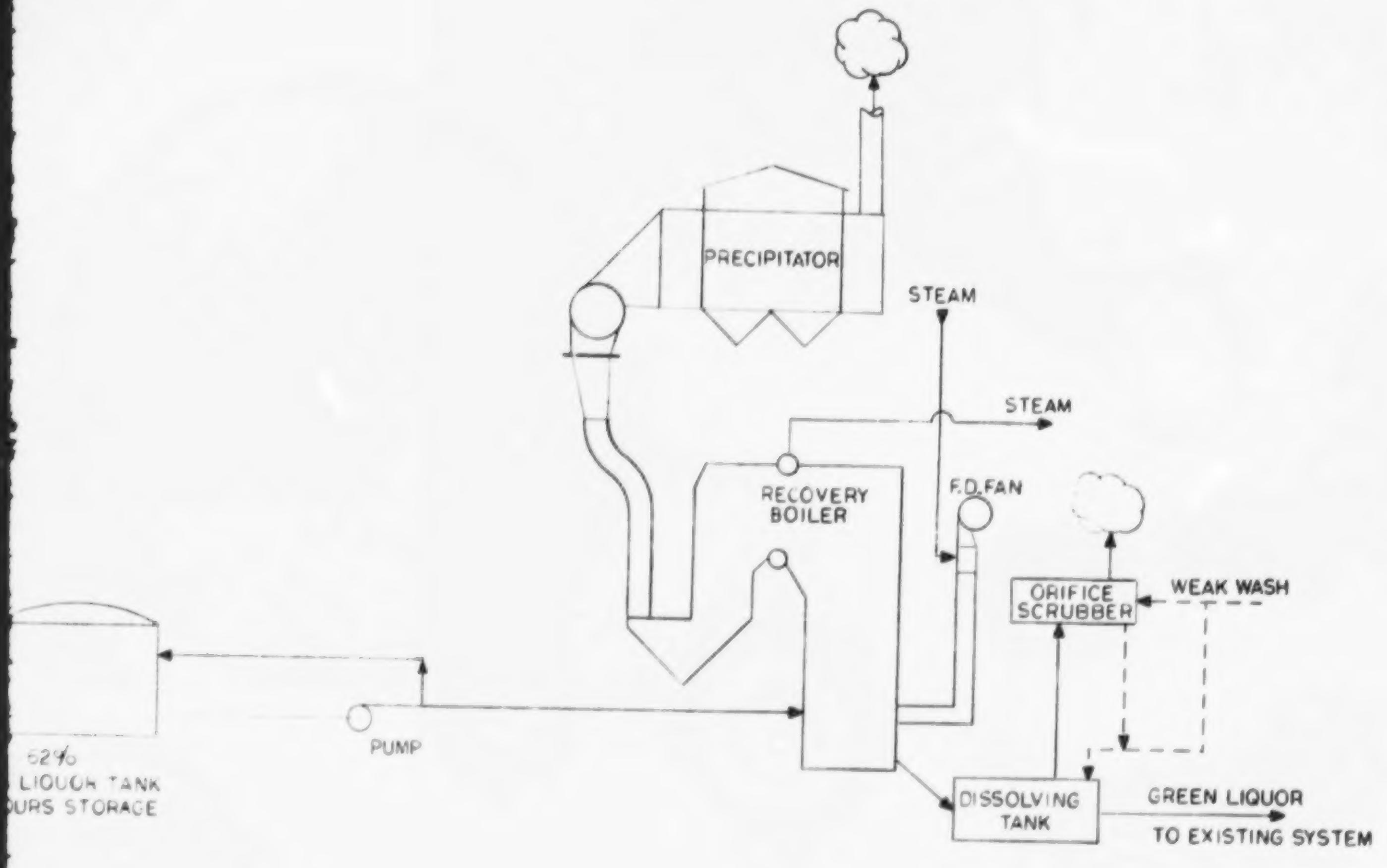
AIR CASCADE EVAPORATOR KRAFT RECOVERY SYSTEM

62%
AIR C

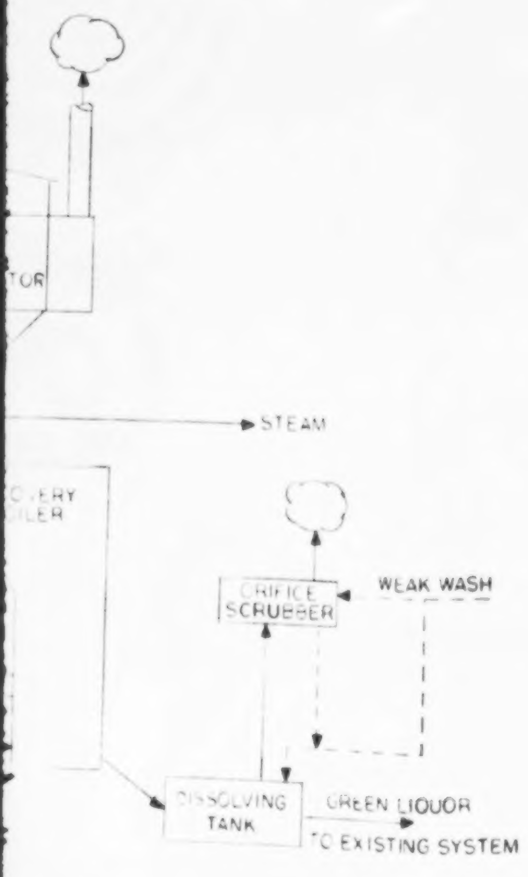
ENV

B

1


PRECIPITATOR
STEAM
STEAM
RECOVERY BOILER
F.D. FAN
ORIFICE SCRUBBER
WEAK WASH
62%
LIQUOR TANK
OURS STORAGE
PUMP
DISSOLVING TANK
GREEN LIQUOR
TO EXISTING SYSTEM

AFT RECOVERY SYSTEM


TOR
STEAM
OVERY
OILER
ORIFICE SCRUBBER
WEAK WASH
DISSOLVING TANK
GREEN LIQUOR
TO EXISTING SYSTEM

SCHEMATIC ARRANGEMENTS 62% BLACK LIQUOR SOLIDS KRAFT RECOVERY SYSTEM AND AIR CASCADE EVAPORATOR KRAFT RECOVERY SYSTEM.	EXHIBIT NO. FIG. 5-44 HJS

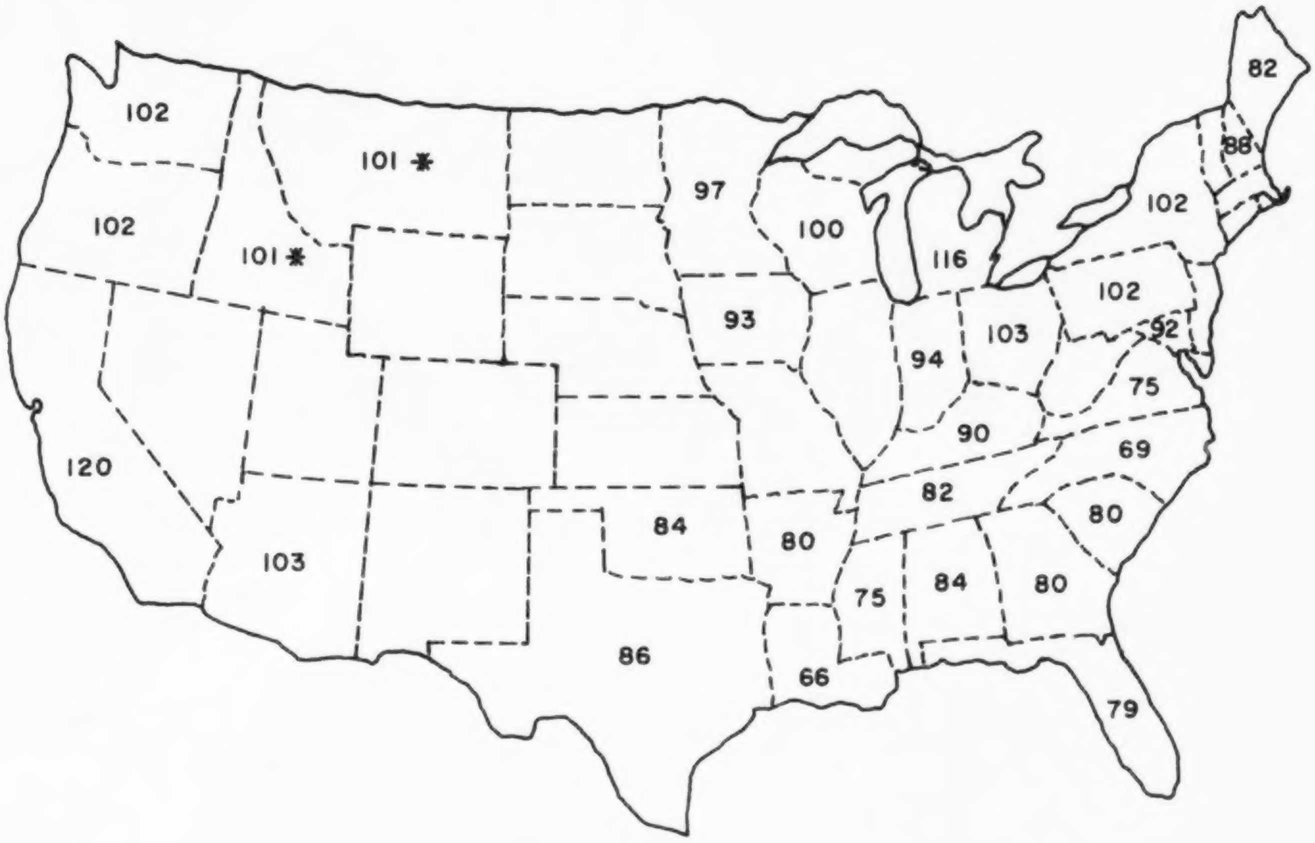
SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

FIGURE 5-45

CONSTRUCTION LABOR COST INDEX


82
88
102
101 *
97
102
100
101 *
116
102
93
103
92
94
75
90
69
120
82
80
84
80
103
75
84
80
86
66
79

THE LABOR INDEX SHOWN BY STATE IS BASED ON AVERAGE OF SELECTED CITY INDICES WITHIN THE STATE.

STATES NOT HAVING PULP MILLS NOT SHOWN.

INDEX DATA FROM "BUILDING CONSTRUCTION COST DATA 1969" 27TH EDITION, ROBERT SNOW MEANS CO., INC. EXCEPT * FROM "DOD COST REVIEW GUIDE, 1970."

INDEX 100 = 1968 U.S. AVERAGE

2. An electrostatic precipitator at specified efficiency located on the building roof. The precipitator would be a filled tile shell unit of two-chamber construction.

3. The recovery building would be enclosed with asbestos or equivalent siding up to approximately the mud drum level. The floors in the building would be of concrete construction.

4. A control room is included for this recovery system. While it may be desirable to include the controls for this unit in a central control room, the capital costs are based on a control room adjacent to this unit.

5. All liquor piping is based on carbon steel material.

6. All equipment which is within the recovery building is included in the estimate. The following additional equipment which is usually located outside the building is also included: (No building enclosures are included for this equipment located outside the recovery building).

    a. Piping tie-ins for green liquor, weak wash, salt cake make-up, high pressure steam, 50 lb. steam, and feedwater and black liquor

    b. Black liquor dump tank

    c. A black liquor evaporator concentrator is included for the systems which require this.

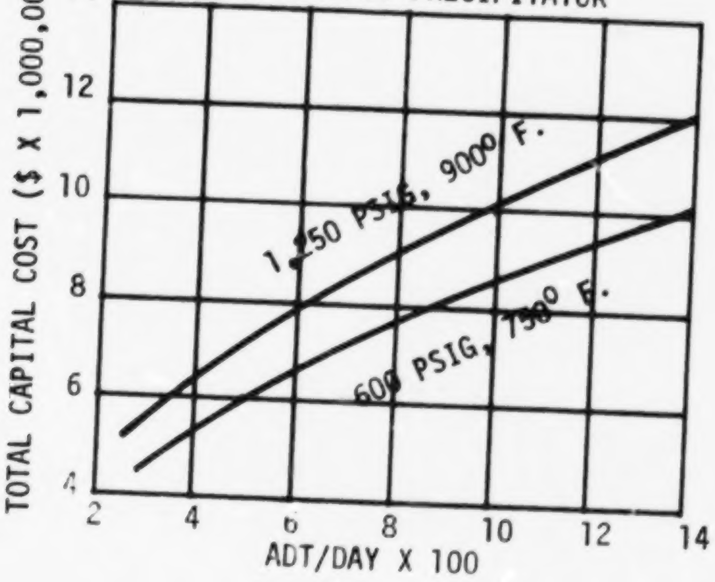
    d. Concentrated black liquor oxidation unit for the recovery system utilizing oxidation

    e. Black liquor storage tank for 62 percent solids and 55 percent solids liquor system; 12-hour storage is included.

Capital cost curves are shown in Figures 5-46 and 5-47.

It should be pointed out that these cost curves do not include any potential credits for better smelt reduction, increased pulp production, increased steam production, or manpower reductions. Depending on individual circumstances, the credits may be a substantial amount.

CONVENTIONAL KRAFT RECOVERY SYSTEM
99.5% EFFICIENCY PRECIPITATOR



CONVENTIONAL KRAFT RECOVERY SYSTEM WITH
CONCENTRATED BLACK LIQUOR OXIDATION 99.5%
EFFICIENCY PRECIPITATOR

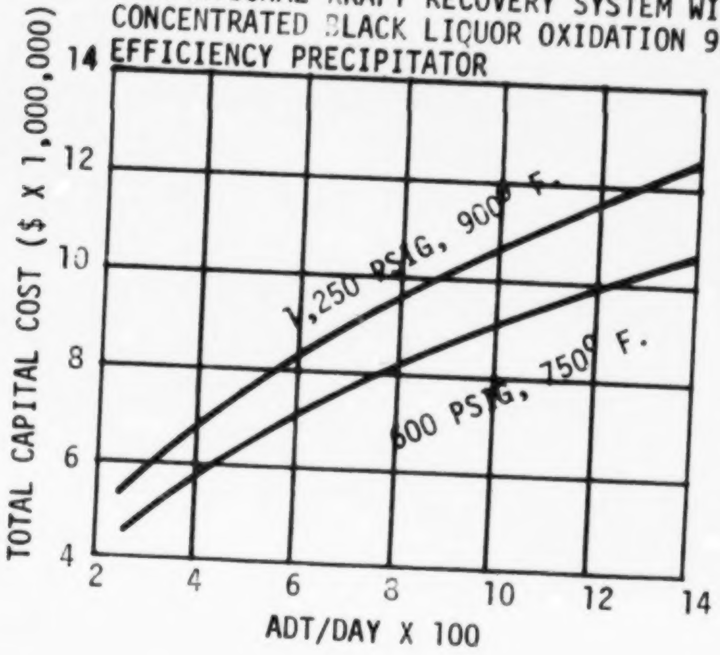
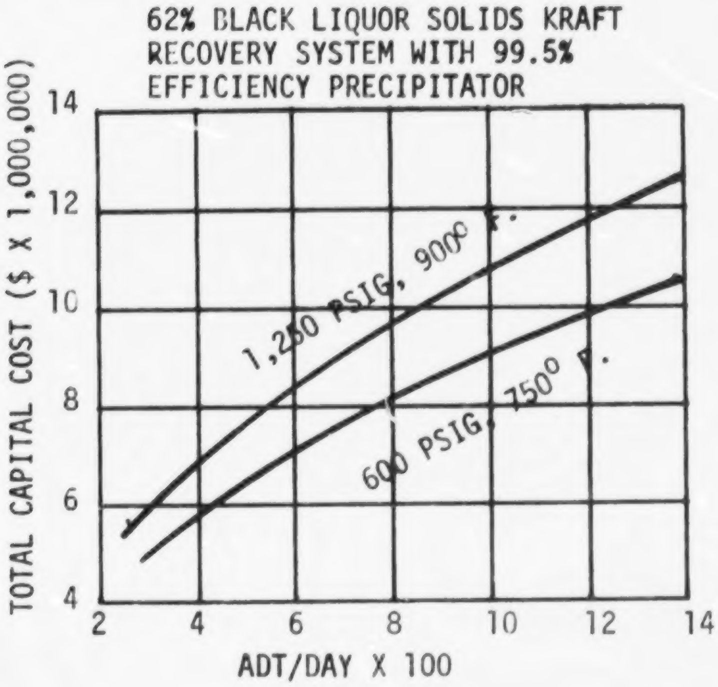
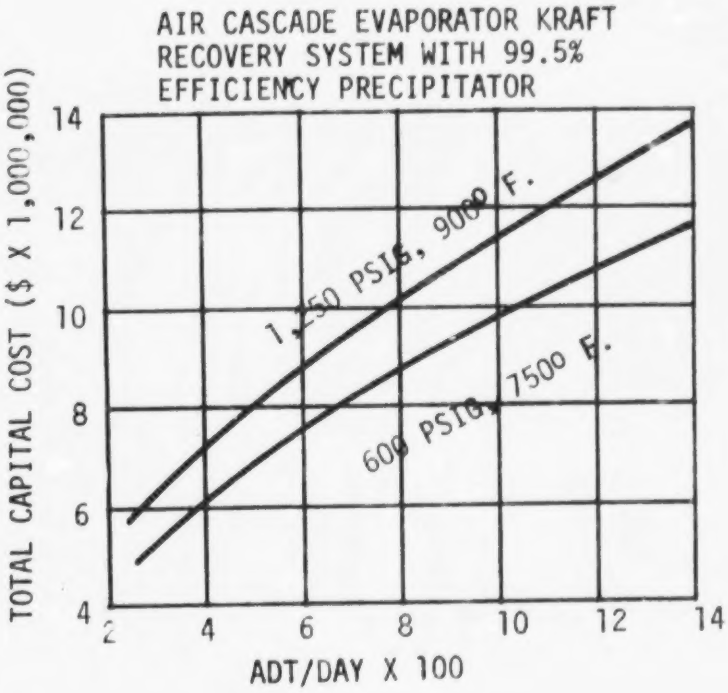


FIG. 5-46 TOTAL CAPITAL COST BASED ON 3,000 LBS. SOLIDS?ADT AND 6,600 BTU/LB. DRY SOLIDS



62% BLACK LIQUOR SOLIDS KRAFT RECOVERY SYSTEM WITH 99.5% EFFICIENCY PRECIPITATOR



AIR CASCADE EVAPORATOR KRAFT RECOVERY SYSTEM WITH 99.5% EFFICIENCY PRECIPITATOR

FIG. 5-47 TOTAL CAPITAL COST BASED ON 3,000 LBS. SOLIDS/ADT AND 6,600 BTU/LB. DRY SOLIDS

5.3.6.2 Smelt Dissolving Tank

Application

The molten inorganic chemicals leave the furnace through smelt spouts to an agitated dissolving tank which contains weak wash liquor obtained from washing lime sludge in the succeeding causticizing operation. After being cooled and dissolved in the weak wash, the smelt solution is referred to as green liquor. Shatter nozzles are used to inject steam and/or recirculated green liquor into the solid smelt stream from the furnace to disperse the molten chemicals and ensure its safe dissolution.

Large quantities of water vapor are released by the green liquor which cools the molten smelt and by the steam from the shatter jets. Particles of smelt and droplets of green liquor become entrained in these vapors and are exhausted through the smelt dissolving tank vent.

The gas vented from the smelt dissolving tank has a temperature of approximately 200°F. and contains traces of organic sulfur compounds and particulate matter with a concentration of 1 - 1.5 grains/SCFD. The particulate matter is mostly sodium sulfide and sodium carbonate.

The majority of mills provide control equipment to minimize the emission of these chemicals and droplets to the atmosphere. Control equipment which is currently in use is as follows:

1. Mesh Pads. Mesh pads are used extensively to collect chemicals in dissolving tank vents. Sprays located above the mesh pads operate periodically to remove the collected chemicals which are returned to the smelt dissolving tank. A fan is not normally required due to the low pressure drop.

2. Packed Towers. Packed towers have been reported at two mills where they operate successfully. One concern with packed towers is the possibility of plugging; however, periodic use of condensate as scrubbing liquid is reported to remove any solids buildup.

3. Orifice Scrubber. This scrubber has been installed at several mills and is reported to be successful where the scrubber is adequately sized. Due to the pressure drop (approximately 8 inches H_2O) of this scrubber, inadequate sizing or improper operation may have a significant effect on emissions (For example, higher than design flows may require by-passing the scrubber.).

Costs

Cyclonic Scrubber. This control method is based on the installation of a single cyclonic scrubber on an elevated floor within the confines of an existing recovery boiler building and adjacent to the existing vent stack. Since the capital cost and operating cost of the cyclonic scrubber is very similar to a packed tower, costs have not been calculated for the cyclonic scrubber. The reader is referred to the costs on the packed tower.

Packed Tower. This control method is based on the installation of a single packed tower on an elevated floor within the confines of an existing recovery boiler building and adjacent to the existing vent stack. In addition, the following auxiliaries are included: an axial flow fan and motor located in the outlet of the tower, a circulation pump and motor, instruments and controls for proper operation of the above, piping for make-up, circulation, and tower effluent to dissolving tank, and revised ductwork, including a by-pass to connect the tower to the existing vent stack. Cost curves are presented in Figure 5-48.

Orifice Scrubber. This control method is based on the installation of a single low energy scrubber on an elevated floor within the confines of an existing recovery boiler building and adjacent to the existing vent stack. In addition, the following auxiliaries are included: an axial flow fan and motor located in the outlet of the tower, a circulation pump and motor, instruments and controls for proper operation of the above, piping for make-up circulation and scrubber effluent to dissolving tank, and revised ductwork including a by-pass to connect the scrubber to the existing vent stack. Cost curves are presented in Figure 5-49.

Mesh Pad. This control method is based on the installation of a mesh pad assembly in an existing recovery boiler vent stack.

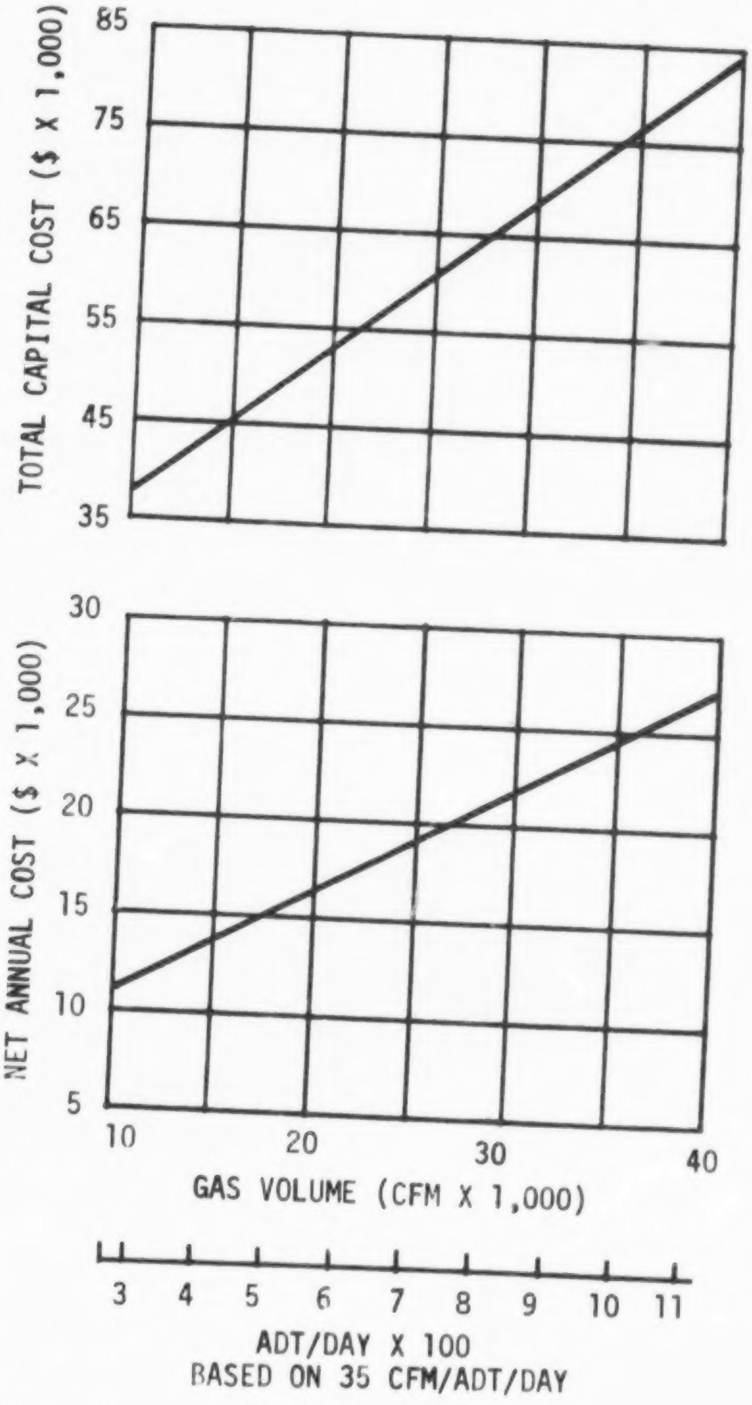


FIG. 5 - 48 CONTROL METHOD COSTS FOR PACKED TOWER ADDED TO SMELT DISSOLVING TANK VENT

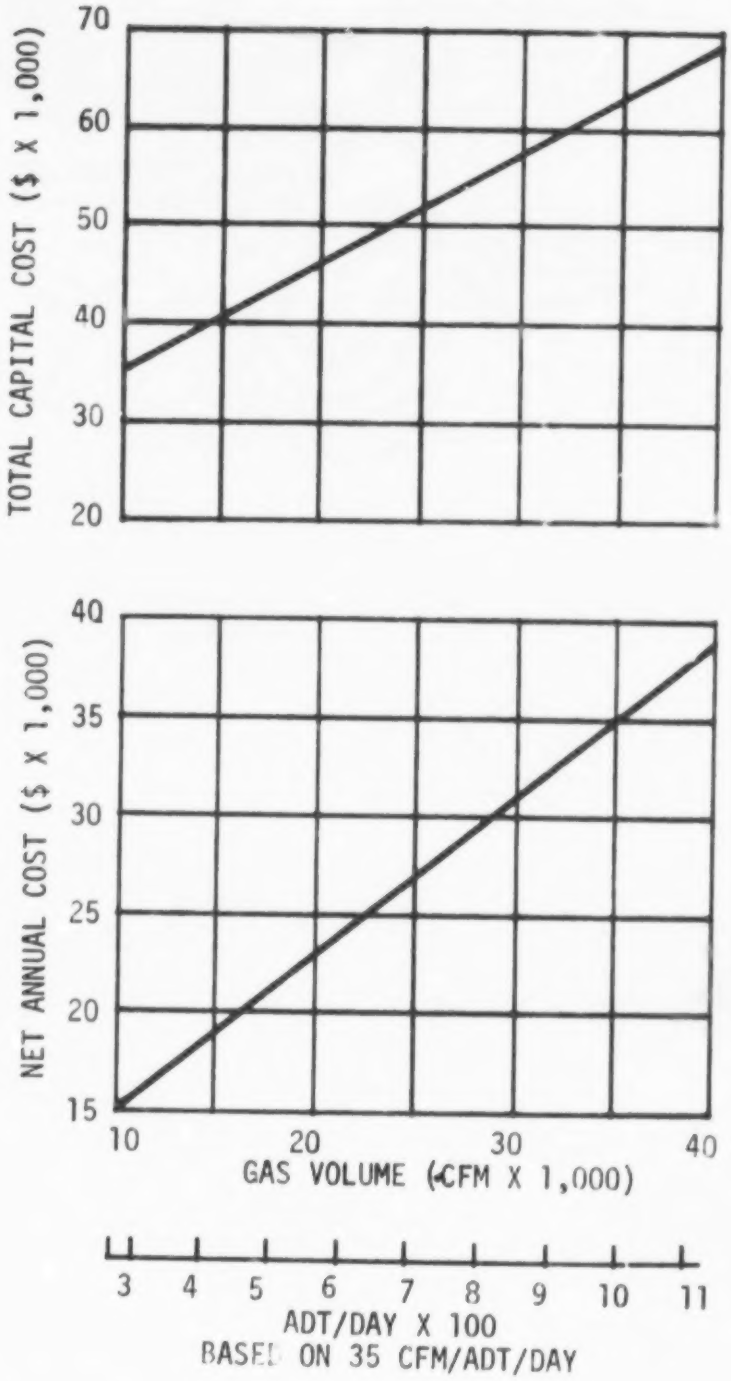


FIG. 5 - 49 CONTROL METHOD COSTS FOR ORIFICE SCRUBBER ADDED TO SMELT DISSOLVING TANK VENT

Allowances have been included for the following: alterations to existing vent stack, normal piping for spray wash, and instruments and controls for proper operation. Cost curves are presented in Figure 5-50.

Effectiveness

Particulate Removal. Efficiencies for the methods considered are as follows:

Control Method	Annual Operating Efficiency
Cyclonic Scrubber	80
Packed Tower	90
Orifice Scrubber	97
Mesh Pad	75

Curves for particulate cost effectiveness are shown in Figures 5-51 and 5-52.

Reduced Sulfur and Sulfur Dioxide Removal. Depending on the scrubbing liquid used, both the packed tower and orifice scrubber would have potential for removing sulfur compounds.

Operation. Of the methods considered, the orifice scrubber should be the least subject to plugging.

Summary. The orifice scrubber and the packed tower are the most effective control methods for either particulate or gaseous pollutant removal. The orifice scrubber is less susceptible to pluggage and has the highest particulate collection efficiency.

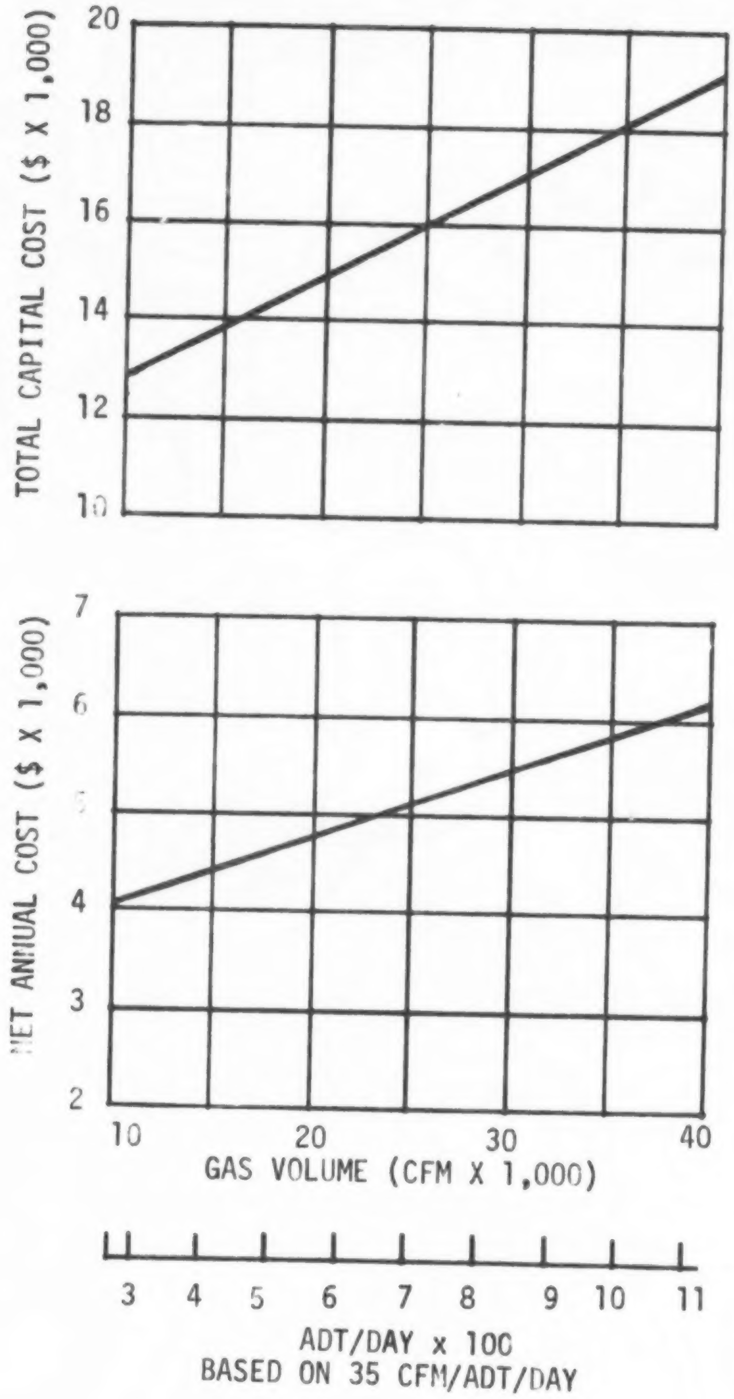
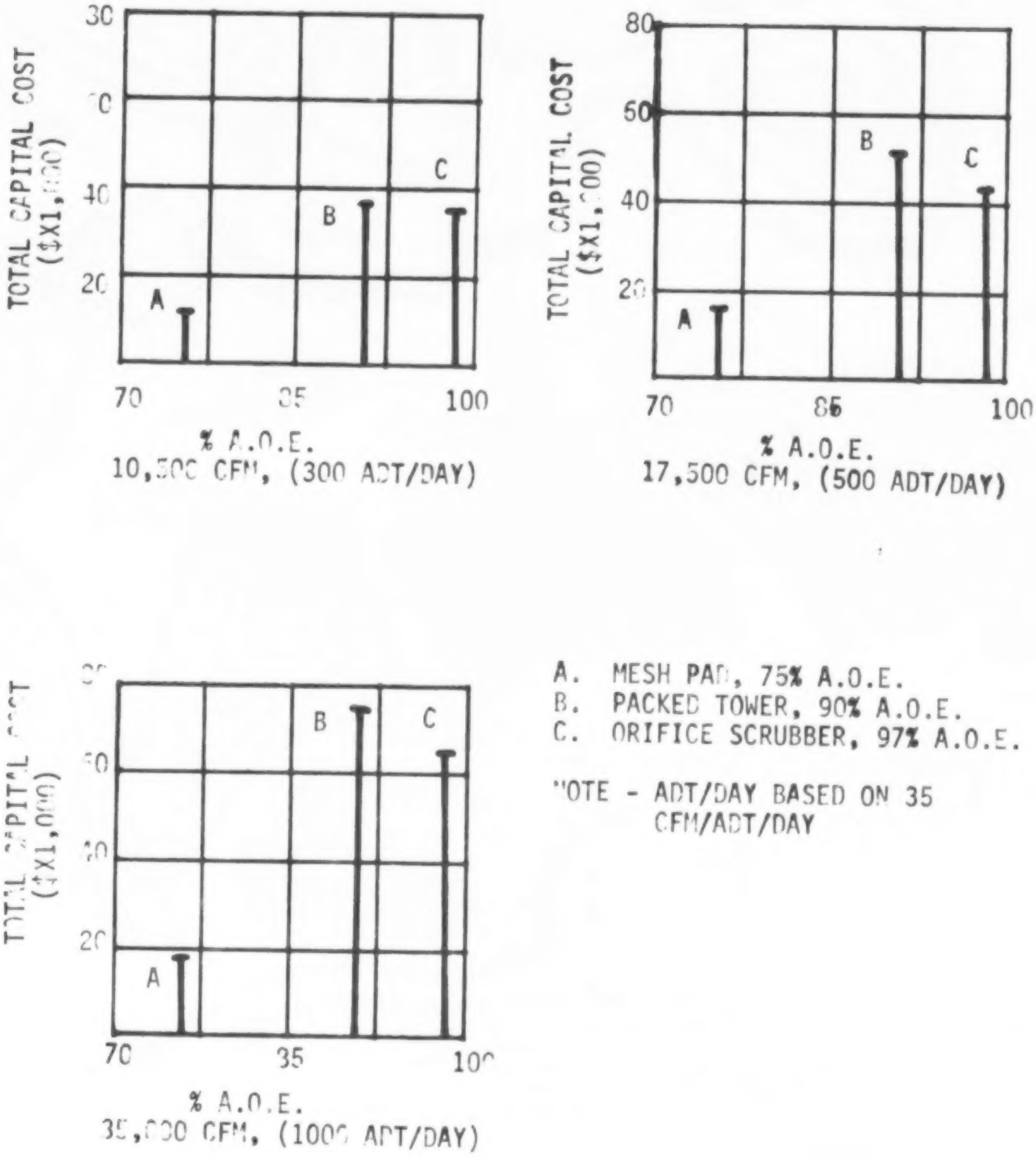


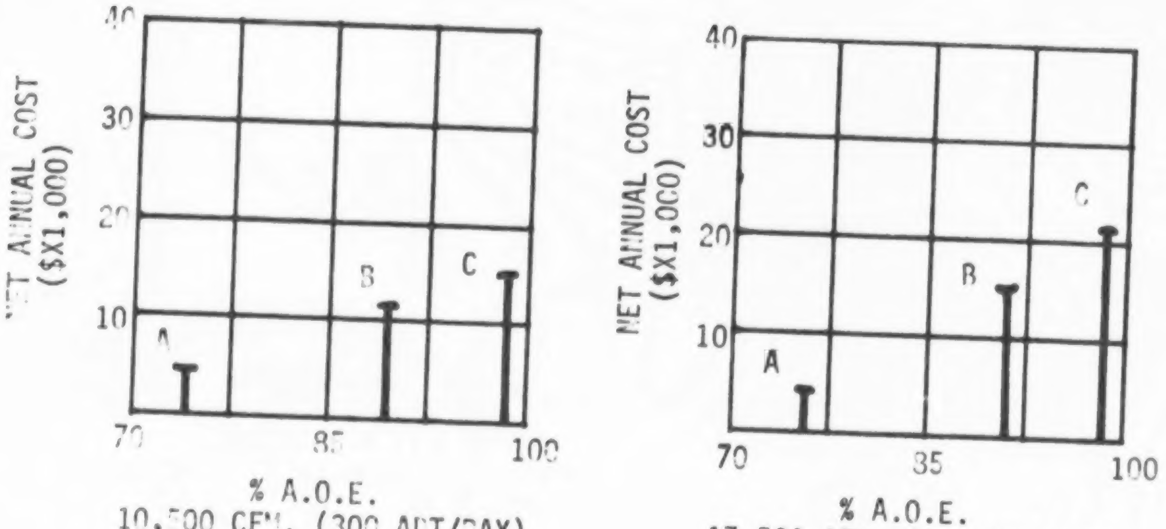
FIG. 5 - 50 CONTROL METHOD COSTS OF MESH PAD ADDED TO SMELT DISSOLVING TANK VENT



A. MESH PAD, 75% A.O.E.
B. PACKED TOWER, 90% A.O.E.
C. ORIFICE SCRUBBER, 97% A.O.E.

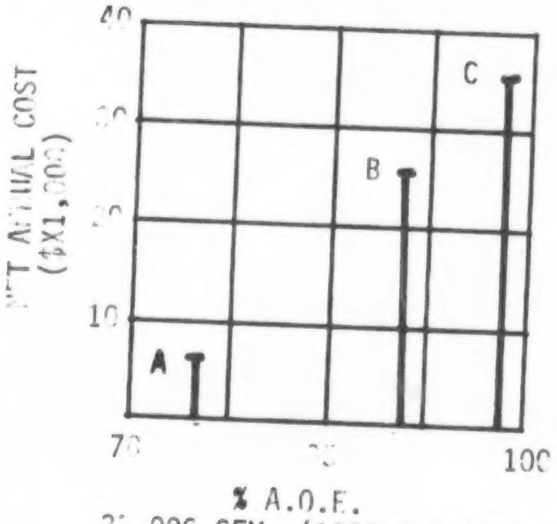
NOTE - ADT/DAY BASED ON 35
CFM/ADT/DAY

FIG.5-51 PARTICULATE COST EFFECTIVENESS (TOTAL CAPITAL COST COMPARISON) FOR CONTROL EQUIPMENT ADDED TO SMELT DISSOLVING TANK VENT - RECOVERY BOILER



10,500 CFM, (300 ADT/DAY)

17,500 CFM, (500 ADT/DAY)



35,000 CFM, (1000 ADT/DAY)

A. MESH PAD, 75% A.O.E.
B. PACKED TOWER, 90% A.O.E.
C. ORIFICE SCRUBBER, 97% A.O.E.

NOTE - ADT/DAY BASED ON 35 CFM/ADT/DAY

FIG.5 - 52 PARTICULATE COST EFFECTIVENESS (NET ANNUAL COST COMPARISON) FOR CONTROL EQUIPMENT ADDED TO SMELT DISSOLVING TANK VENT - RECOVERY BOILER

5.3.6.3 Digester Relief and Blow, Multiple Effect Evaporators

Application

The gases leaving the batch digester blow tank, turpentine condensers, and multiple effect evaporators all contain high concentrations of organic sulfur compounds. Despite the fact that the gas volumes are rather small, the high concentration of odorous compounds makes these vents major contributions to the odor level in the mill. Because a continuous digester blows into a blow tank at a temperature below the flash point there are no gases venting from a continuous digester blow tank. Few mills have control equipment on these sources. With the present emphasis on odor control, much attention lately has been given to these sources. Depending on local conditions, these sources can be treated independently or commonly. The odor removal problem is the same in both cases. Methods which are presently in use are as follows:

Chlorination. This system consists of the equipment necessary to collect and chemically oxidize pulp mill noncondensible gases as shown in Figure 5-53.

The digester relief and blow gases and gases emerging from the evaporator hot well vent and from the turpentine condenser and decanter vents are exposed to chlorination stage washer effluent from the bleach plant, either in the dropleg of the washer or in a scrubber. The dimethyl sulfide (RSR) is absorbed and oxidized to sulfone. The dimethyl disulfide (RSSR) present in the stream of gas is absorbed and oxidized to methyl sulfonyl chloride. This compound consumes a large proportion of the chlorine and may not be fully absorbed and oxidized.

A portion of the required chlorine is usually available in the chlorination stage washer effluent from the bleach plant. Supplemental chlorine may be required. In an unbleached pulp mill, chlorine gas has to be supplied from an outside source. There is a system available that scrubs the gases with chlorine and caustic. This system also treats the condensate by air stripping.

Combustion. This system consists of the equipment necessary to collect and thermally oxidize the sulfur bearing compounds in pulp mill noncondensibles to sulfur dioxide by incineration in the lime kiln or a separate incinerator if the distance from the source to the lime kiln makes the installation cost prohibitive. Sulfur dioxide is considered less objectionable and more easily removed than the compounds from which it was formed. Figure 5-54 shows the arrangement of this system.

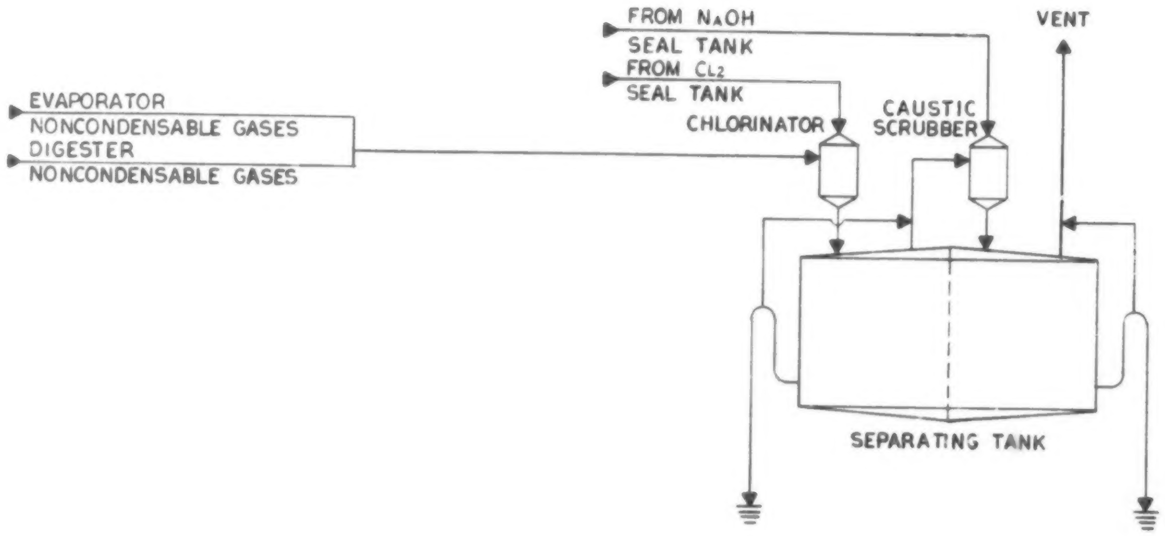
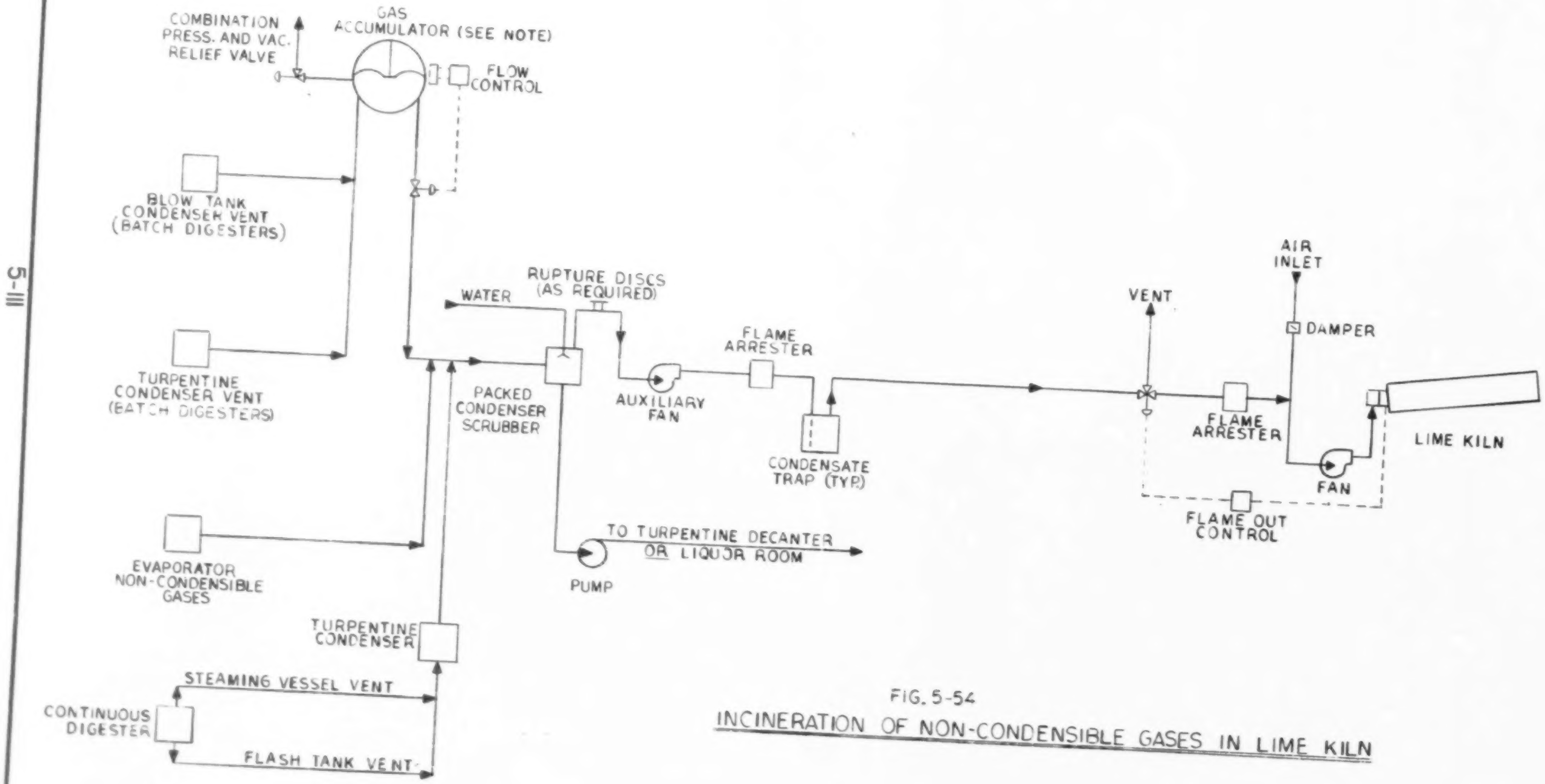

FROM NaOH
SEAL TANK
FROM Cl2
SEAL TANK
VENT
EVAPORATOR
NONCONDENSABLE GASES
DIGESTER
NONCONDENSABLE GASES
CHLORINATOR
CAUSTIC
SCRUBBER
SEPARATING TANK

FIG. 5-53
Chlorination System


COMBINATION
PRESS. AND VAC.
RELIEF VALVE
GAS
ACCUMULATOR (SEE NOTE)
FLOW
CONTROL
BLOW TANK
CONDENSER VENT
(BATCH DIGESTERS)
TURPENTINE
CONDENSER VENT
(BATCH DIGESTERS)
EVAPORATOR
NON-CONDENSIBLE
GASES
TURPENTINE
CONDENSER
STEAMING VESSEL VENT
CONTINUOUS
DIGESTER
FLASH TANK VENT
WATER
RUPTURE DISCS
(AS REQUIRED)
PACKED
CONDENSER
SCRUBBER
AUXILIARY
FAN
FLAME
ARRESTER
CONDENSATE
TRAP (TYP.)
PUMP
TO TURPENTINE DECANTER
OR LIQUOR ROOM
VENT
FLAME
ARRESTER
AIR
INLET
DAMPER
FAN
FLAME OUT
CONTROL
LIME KILN

NOTE: GAS ACCUMULATOR IS REQUIRED FOR COLLECTING AND STORING GASES DURING PEAK FLOWS OF A BATCH DIGESTER SYSTEM ONLY

FIG. 5-54

INCINERATION OF NON-CONDENSIBLE GASES IN LIME KILN

In order to accomplish this conversion, the gases must be collected and delivered to the kiln at a constant rate of flow. The noncondensible portion of the digester blow gases discharges periodically from the blow heat condenser vent. These gases are piped to a gas accumulator having sufficient volume to allow collection of the gases as they occur while discharging at a constant rate. This accumulator is required only for use with batch digesters. The noncondensible relief gases from the evaporator hot well vent occur at essentially a steady rate and need not be collected in a storage vessel and can be sent directly to the kiln.

All of the pulp mill noncondensible gases are directed to a packed condenser-scrubber for the removal of turpentine mist from the gas stream. Turpentine mist may interfere with the flame out control through development of a false signal. From the scrubber the gases are directed to the suction of the kiln fan, which is located near the kiln. At this point the gases are diluted beyond the limits of inflammability and introduced to the firing end of the kiln. At the two thousand degree temperature usually encountered in the lime kiln, the conversion takes place rapidly and completely.

Safety measures in the form of flame arrestors are required to avoid explosive conditions ahead of the lime kiln.

Caustic Scrubber. The scrubber is used to remove H_2S and mercaptans by absorbing the gas in white liquor. (See Figures 5-55 and 5-56).

The method of collection of the pulp mill noncondensibles in this system is similar to the method used for combustion previously discussed. A notable exception is that the gas holder for use with batch digesters can be omitted. This is made possible by using a scrubber sized to take the surges of gas that occur. The volume supplied to the scrubber is actually constant, with the gas to air ratio changing when a digester is blown.

Catalytic Oxidation. Catalytic methods of oxidizing noncondensible gases have been tried. So far their use has not proven applicable to this source due to severe operating problems and excessively high maintenance costs (12).

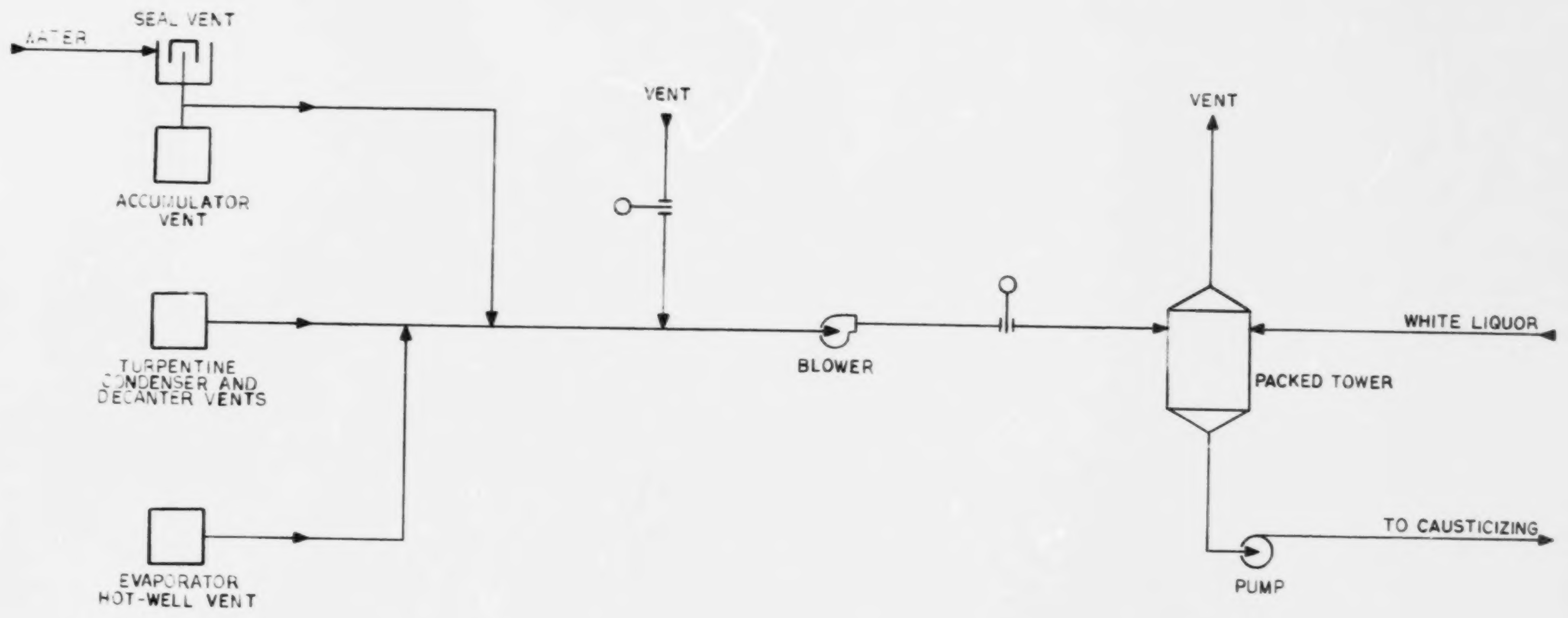

WATER
SEAL VENT
ACCUMULATOR VENT
TURPENTINE CONDENSER AND DECANTER VENTS
EVAPORATOR HOT-WELL VENT
VENT
BLOWER
VENT
PACKED TOWER
WHITE LIQUOR
PUMP
TO CAUSTICIZING

FIG. 5-55

CAUSTIC SCRUBBER - BATCH DIGESTERS

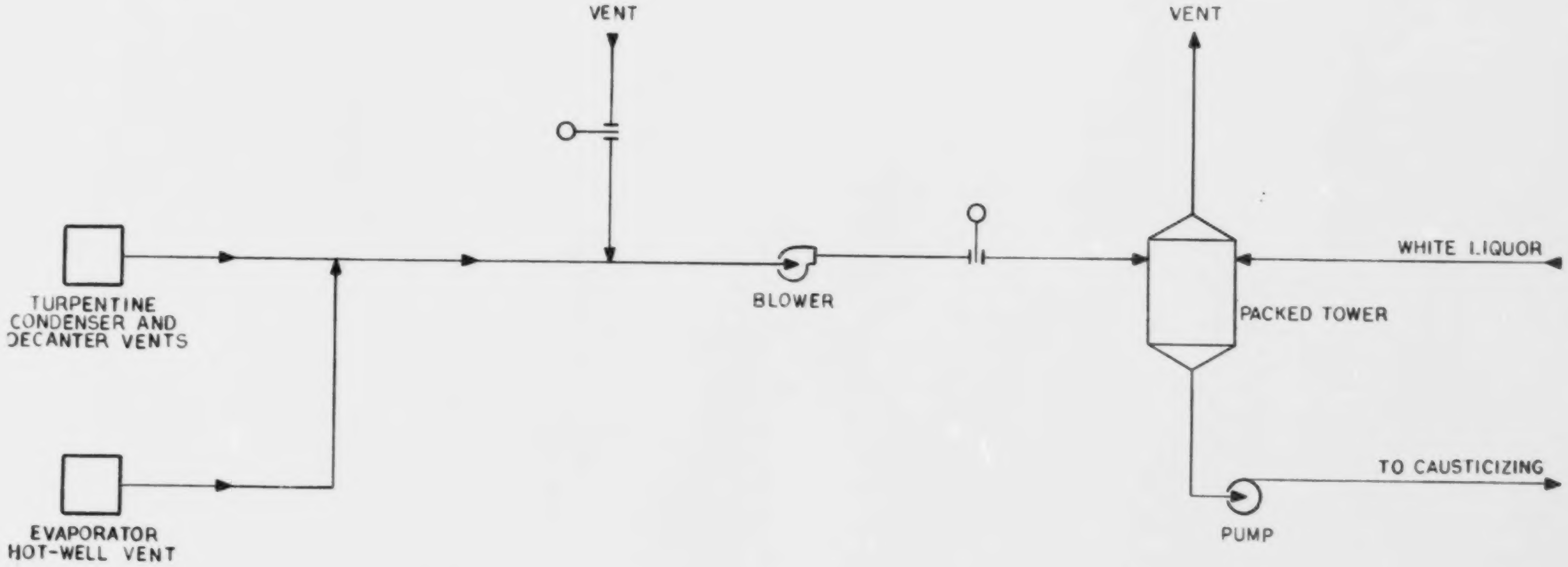

VENT
VENT
TURPENTINE CONDENSER AND DECANTER VENTS
EVAPORATOR HOT-WELL VENT
BLOWER
PACKED TOWER
WHITE LIQUOR
TO CAUSTICIZING
PUMP

FIG. 5-56

CAUSTIC SCRUBBER - CONTINUOUS DIGESTERS

Costs

Following are cost curves for various systems showing the total installed cost and the annual operating cost. (Figures 5-57 through 5-61).

Effectiveness

Particulates. There are no particulate emissions from this source.

Reduced Sulfur Removal. Of the methods discussed, combustion could be expected to have an effectiveness approaching 100 percent removal. Chlorination has almost no effect on H_2S, about 98 percent reduction of mercaptan and more than a 50 percent reduction of sulfide. (Phillips said quite a reduction in Hanible Island report).

Sulfur Dioxide Removal. The thermal oxidation of reduced sulfur compounds results in the formation of sulfur oxides. However, limited information developed in field studies by NCASI (19) engineers has failed to show significant amounts of sulfur oxides in the kiln gases when incinerating the noncondensibles. The phenomena which reduce these sulfur oxides is unknown; however, it is possible that the alkaline lime material in the kiln bed provides an efficient removal mechanism for the acidic sulfur oxides formed.

The relative levels of sulfur oxides for these control methods are unknown. However, there does not appear to be a significant difference between these methods.

Operation. All of the methods except catalytic oxidation have proved to be practical from an operational point of view.

Summary. At the present time, combustion of noncondensible gases appears to be favored by mills in the United States because of the 99 plus percent removal of total reduced sulfur compounds.

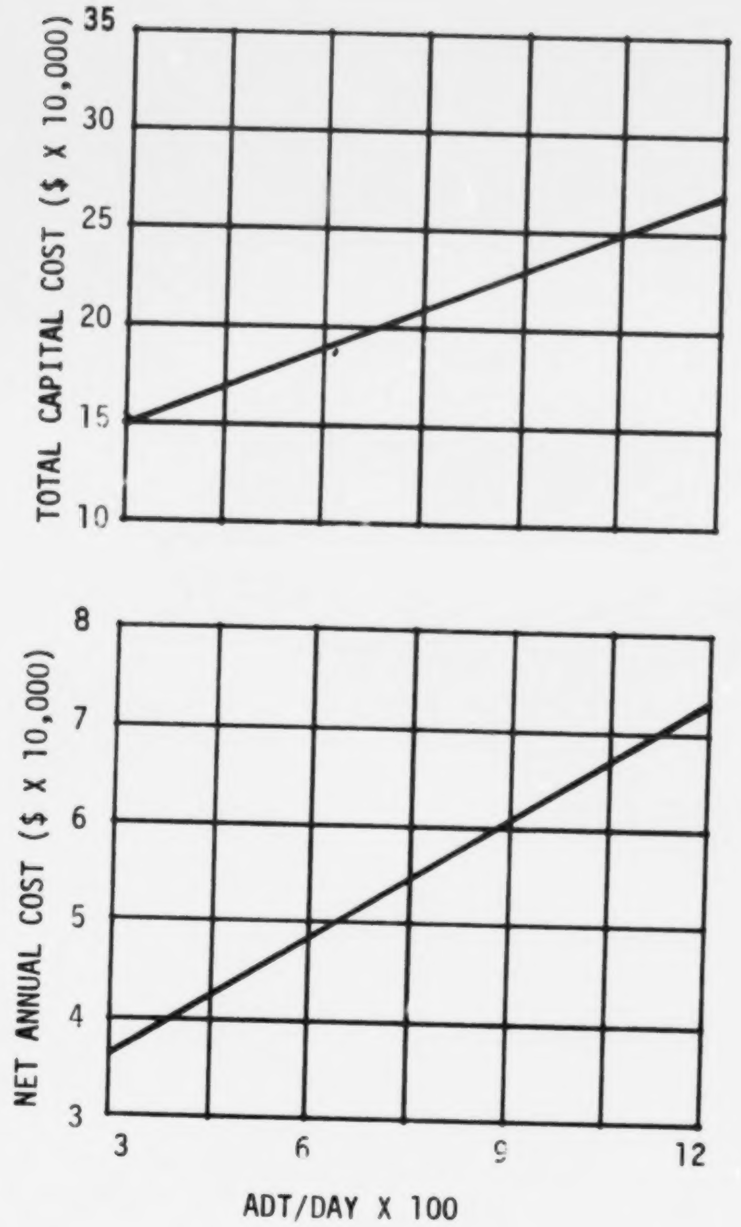


FIG.5 - 57 CONTROL METHOD COSTS FOR CHLORINATION SYSTEM APPLIED TO NON-CONDENSIBLE GASES

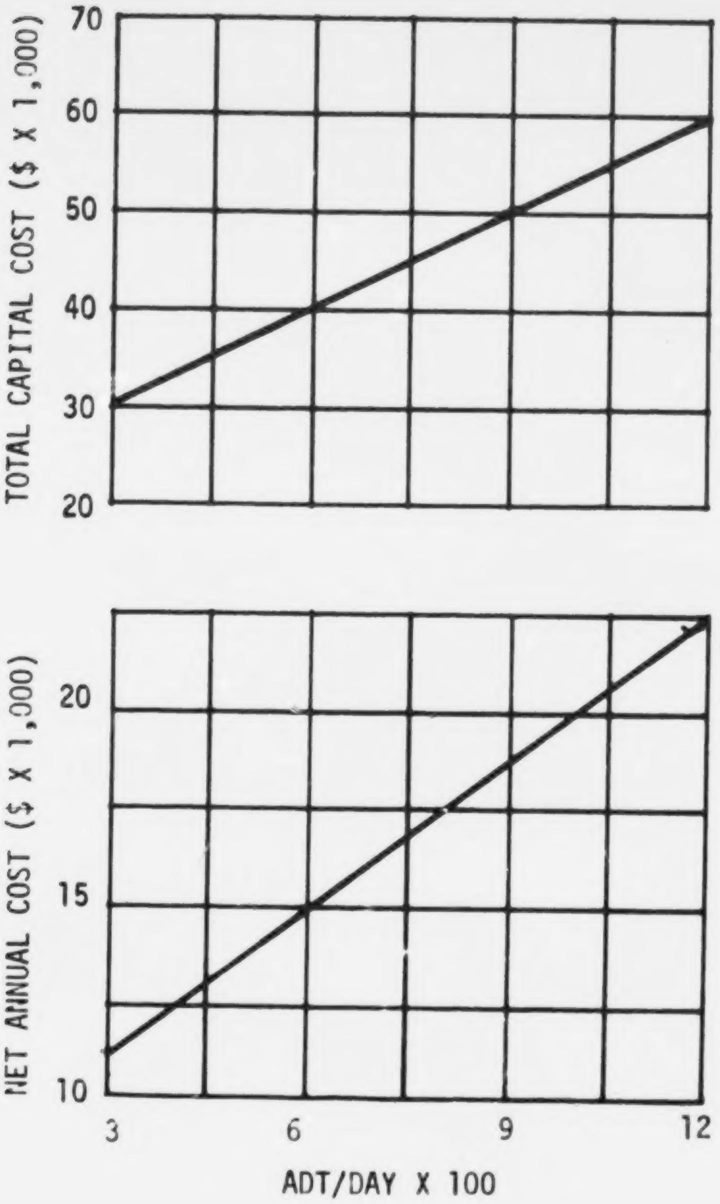


FIG.5-58 CONTROL METHOD COSTS FOR INCINERATION OF NON-CONDENSIBLE GASES IN LIME KILN - CONTINUOUS DIGESTERS

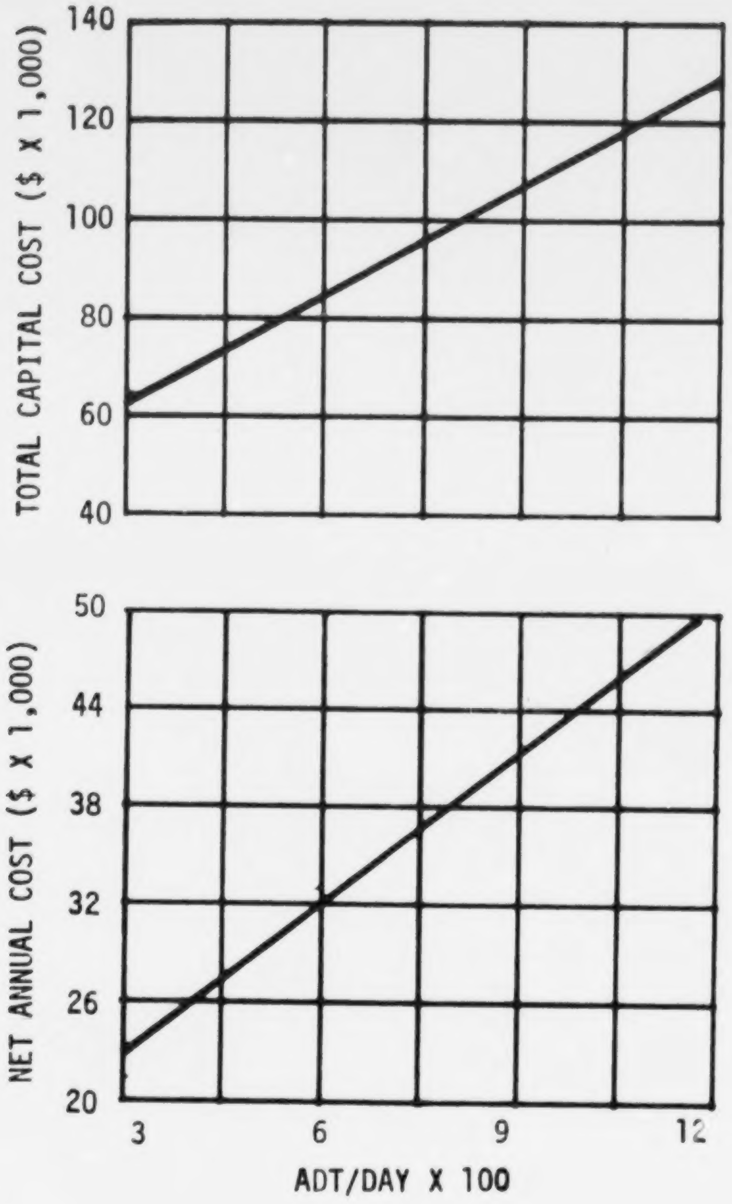


FIG.5-59 CONTROL METHOD COSTS FOR INCINERATION OF NON-CONDENSIBLE GASES IN LIME KILN - BATCH DIGESTERS

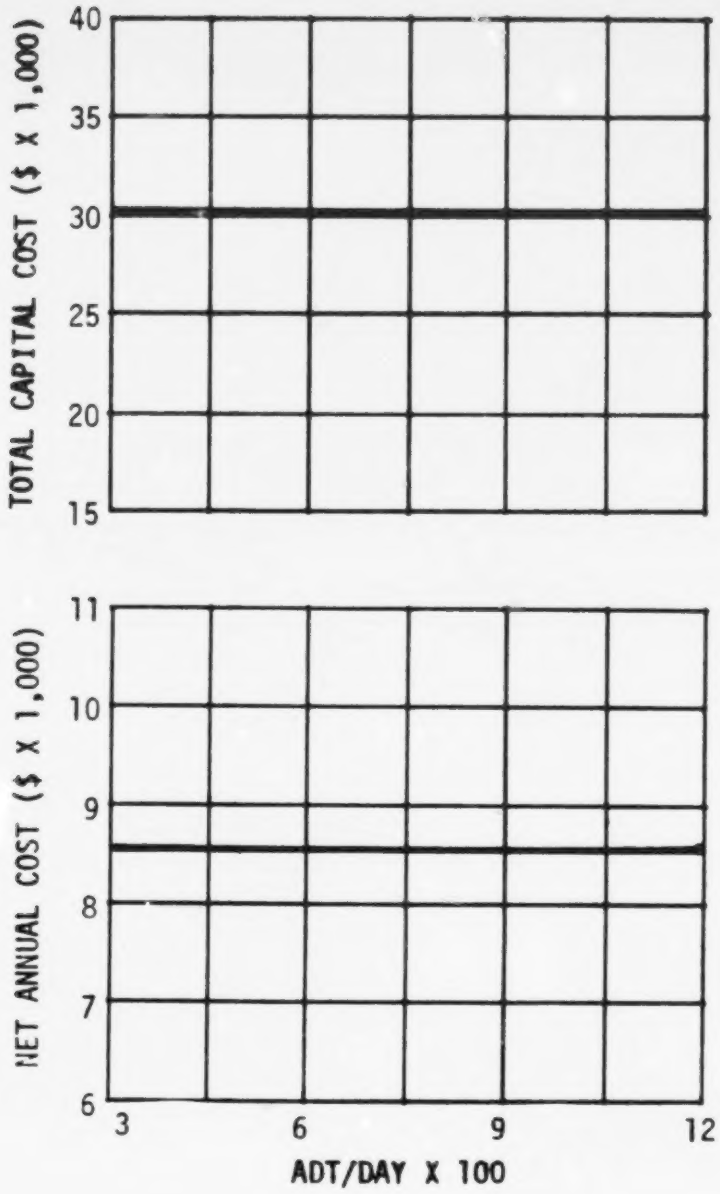


FIG. 5-60 CONTROL METHOD COSTS FOR CAUSTIC SCRUBBER APPLIED TO NON-CONDENSABLE GASES - CONTINUOUS DIGESTERS

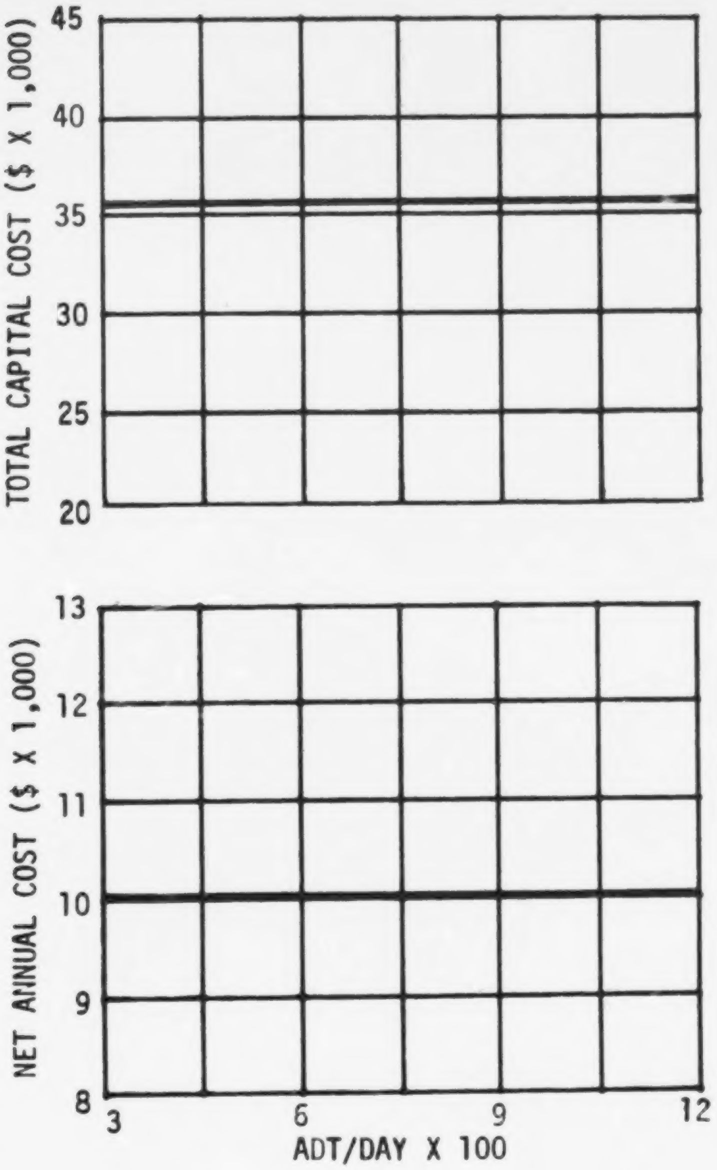


FIG. 5-61 CONTROL METHOD COSTS FOR CAUSTIC SCRUBBER SPPLIED TO NON-CONDENSABLE GASES - BATCH DIGESTERS

5.3.6.4 Lime Kiln

Application

This control method is based upon the addition of a high energy Venturi scrubber to replace existing 80 percent AOE cyclonic type dust collectors. The system is to be arranged such that change over to a new system may be made with a minimum of lost production.

As a matter of convenience and water conservation, most kraft mills use contaminated condensate as a scrubbing liquid in the kiln scrubber. The condensate system is backed up with a fresh water supply. During time of chemical unbalance or lack of supply of condensate, fresh water is substituted as a scrubbing medium.

Any mercaptans which are emitted from the kiln stack come from the contaminated condensate used as a scrubbing liquid and makeup wash water in this system. Substitution of fresh water in these systems would eliminate emission of mercaptans from the kiln stack.

This substitution of fresh water as a scrubbing medium appears simple and relatively inexpensive. However, it must be emphasized that the lime kiln and causticizing areas are the prime users of contaminated condensate generated in other areas of the mill. If fresh water is used in place of contaminated condensate, the load to the waste treatment system will be increased by the amount of water substituted.

The emissions from incineration of noncondensible sulfur bearing compounds in the lime kiln are not being considered in this particular control method.

Costs

The capital costs are based upon purchase of equipment and installation of the new scrubbing system while the existing system is in operation. Items included are as follows:

1. New Induced Draft Fan with Drive
2. New Scrubber, Mist Eliminator and Stack
3. New Pumps with Drives
4. Modification to Existing Hot Gas Ductwork

Also included in the costs are new foundations and structures required to support the equipment and demolition cost for removal of the existing scrubbing system, once the new system is placed in operation. Cost curves are presented in Figures 5-62 and 5-63.

Effectiveness

A single method is considered in this study of replacing a low efficiency cyclonic type dust collector with a high energy Venturi scrubber. The new method is to have a 99.0 or 99.9 percent lime solids collection efficiency. AOE and design efficiencies are considered to be the same.

1. The new system will be to replace an existing and operating inefficient system.

2. The efficiency of solids collection of high energy scrubber is a direct factor of the pressure drop taken across the throat of the scrubber. Because of this item the capital cost between a 99.0 percent and 99.9 percent efficiency would be very similar. Equipment sizing would be essentially duplicate in either case and the only variable would be a higher horsepower motor for driving the induced draft fan. Variables in operating cost are shown on the cost curves.

3. High efficiency scrubbers are important to the pulping industry for recovery of soda. Volatile soda compounds in the kiln exhaust gases are more difficult to recover than solids but are also more valuable. Equipment manufacturers presently estimate that a scrubber with 99.9 percent lime solids collection efficiency will recover 90 to 95+ percent of the soda fume in the kiln exhaust gases. 99.0 percent lime solids collection should be equivalent to approximately 70 percent collection of soda fume.

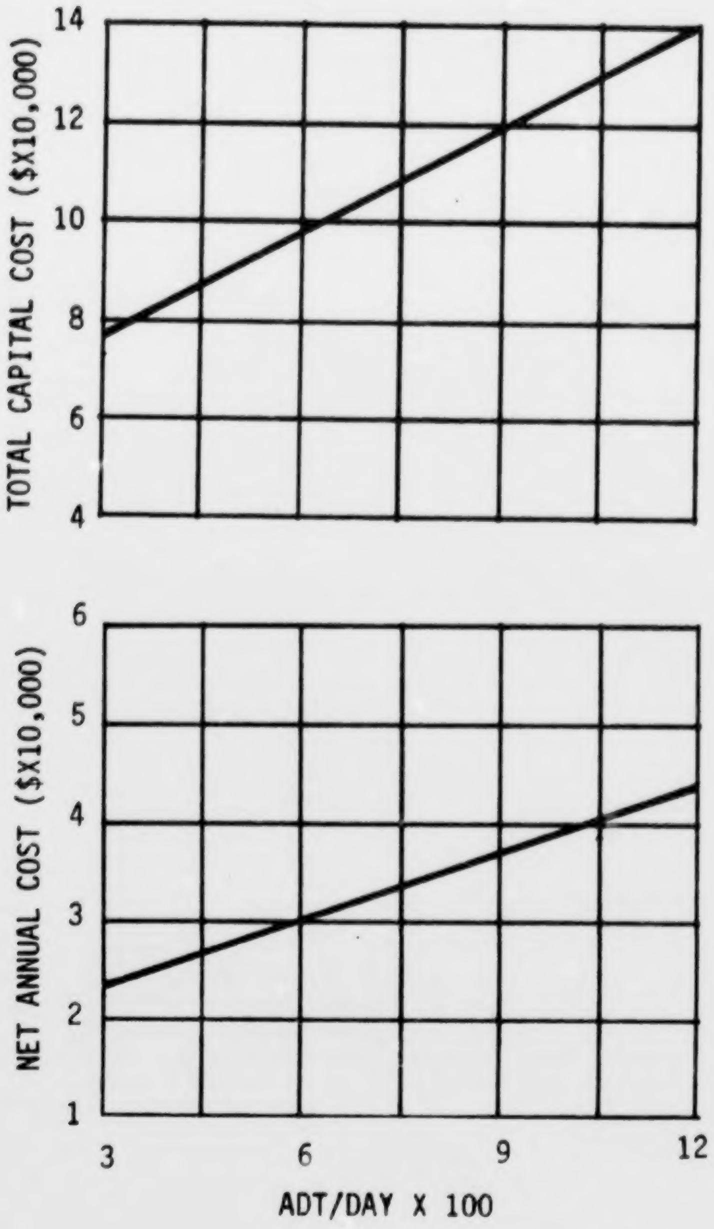


FIG. 5-62 CONTROL METHOD COSTS FOR FRESH WATER VENTURI ADDED TO LIME KILN ---99.0% LIME SOLIDS COLLECTION

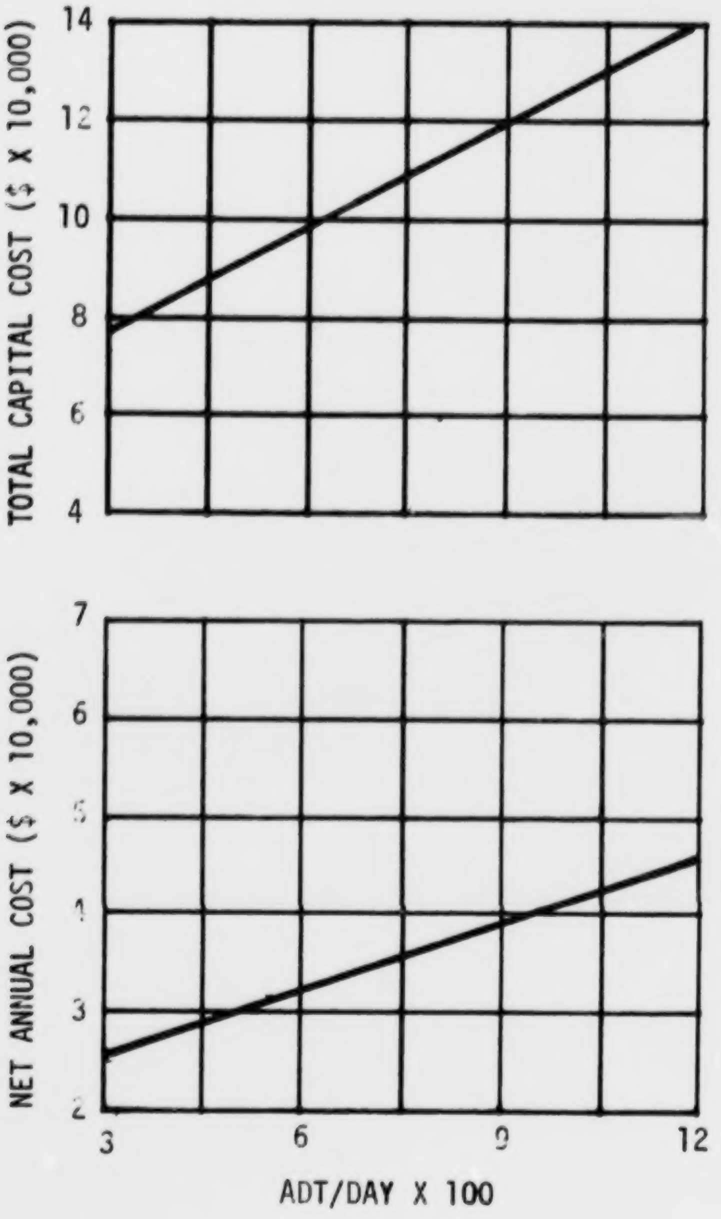


FIG. 5-63 CONTROL METHOD COSTS FOR FRESH WATER VENTURI ADDED TO LIME KILN ---99.9% LIME SOLIDS COLLECTION

5.3.6.5 Lime Slaker

Application

Two control methods are considered for this source of emissions, addition of a cyclonic scrubber and addition of mesh pads. Each application is predicated on installation of a control device on an existing slaker which has no control equipment. Flows from slaker stacks are assumed to be saturated air containing calcium oxide and inert particulates.

Costs

The capital costs are based on installing and operating new control equipment. Demolition of existing vent stacks is included. The estimated systems consist of the following:

1. Figure 5-64

    New Cyclonic Scrubber

    Induced Draft Fan with Drive

    Liquid Supply Pump with Drive

    Vent Stack and Ductwork

    Structural Steel Supports, Foundations, Piping, Controls and Wiring

2. Figure 5-65

    Mesh Pad Mist Eliminator with Housing

    Indusced Draft Fan with Drive

    Liquid Supply Pump with Drive

    Vent Stack

    Structural Steel Supports, Foundations, Piping, Controls and Wiring

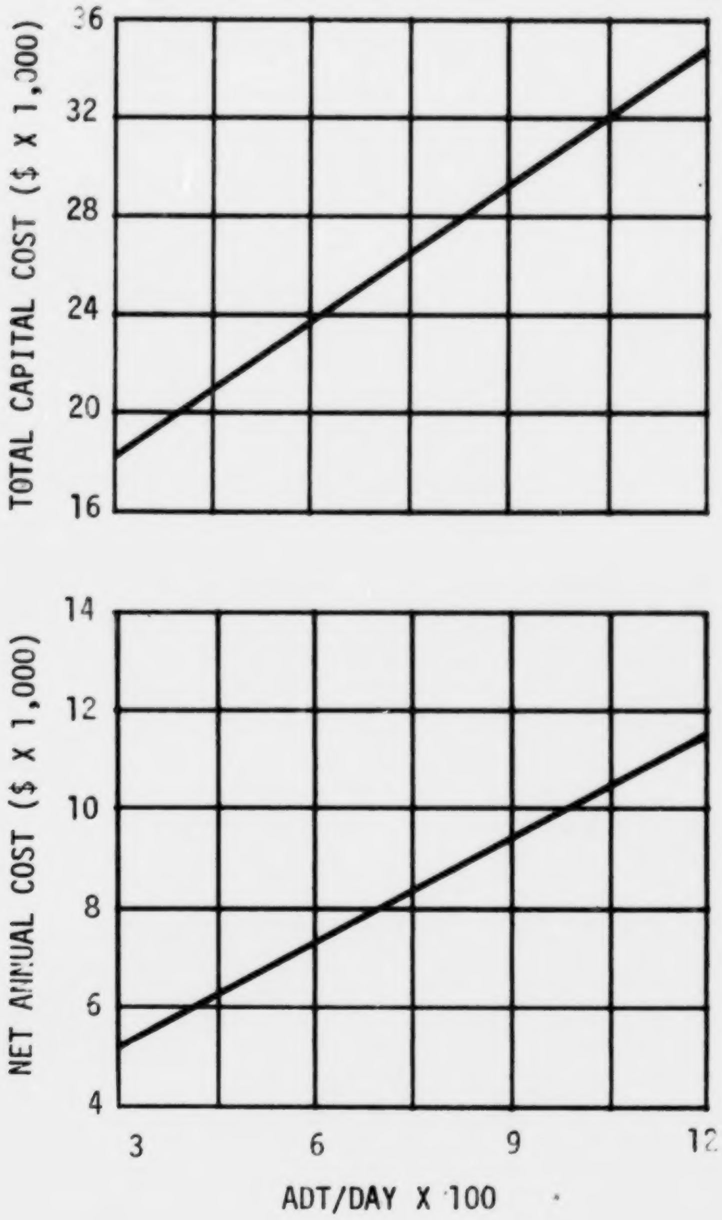


FIG. 5-64 CONTROL METHOD COSTS FOR CYCLONIC SCRUBBER ADDED TO SLAKER

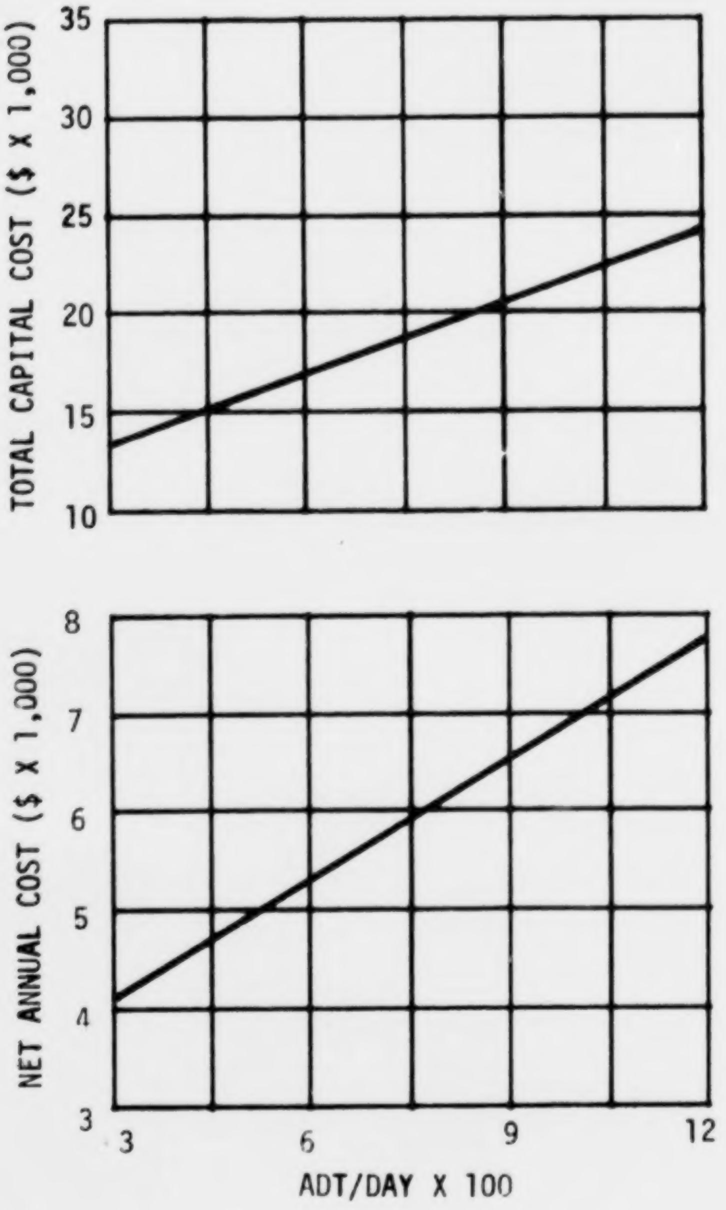


FIG. 5-65 CONTROL METHOD COSTS FOR MESH PAD ADDED TO SLAKER

Effectiveness

No measure of the actual effectiveness of either control method has been developed to date. Further, no accurate measurement of the quantity or quality of pollutants from lime slaker vent stacks has been made. The volume and nature of these materials have been assumed from knowledge of the chemical reaction taking place and from laboratory data.

5.3.6.6 Power and Combination Boilers, General

The emission control equipment used in the steam power plants has been almost exclusively limited to removal of particulate matter. Lately efforts have been made to find ways and means to remove sulfur from the fuels and alternatively sulfur dioxide from the flue gases. As yet no commercially economic and feasible method is in operation, but a number of promising pilot studies have been made. The recognition of nitrogen oxides as undesirable compounds in the flue gases is relatively new and efforts in that direction have only begun.

Power and combination boilers in the pulp and paper industry incorporate a wide variety of firing equipment. Some of these types are:

Oil and Gas Burners

Pulverized Coal

Spreader Stokers with Traveling Grates

Spreader Stokers with Vibrating Grates

Spreader Stokers with Water Cooled Grates

Spreader Stokers with Stationary Grates

Spreader Stokers with Dump Grates

Dutch Ovens

Aside from coal the main fuels are oil, gas, and bark. Today the burning of bark for steam generation is confined to locations where bark is available as a waste or by-product. In producing lumber half of the wood in the log is often discarded as saw-dust, bark, shavings, slabs and ends, all of which can be used as fuel. New methods in barking and pulping, however, are continuously reducing the quantity of waste wood available.

The devices commonly used for removal of particulate matter in flue gases of steam boilers are centrifugal collectors and electrostatic precipitators. The following list indicates the type of dust collectors normally used for various types of fuel.

	Mechanical	Electrostatic Precipitators
Coal (pulverized)	X	X
#6 Oil	X	Has been used but is not common
Gas	-	-
Bark	X	
Bark + Coal	X	
Bark + Oil	X	
Bark + Gas	X	

Fly ash may be approximately classified according to type of firing as shown in Table 5-1. Fly ash from pulverized coal fired boilers is generally quite fine--about 70 percent (or more) samller than 325 mesh. It is usually light gray in color as it contains only a small amount of unburned carbon. The samll particles, having a large surface area, will impart a definite color to the stack gases. This may, under certain light conditions, indicate that a sizable amount of fly ash is being discharged into the atmosphere. Fly ash from stoker fired boilers on the other hand is much coarser and has, since it contains larger quantities of unburnt carbon, a dark gray to black color. This fly ash does not discolor the flue gas in the same manner as described above. A clean appearing stack discharge from a stoker fired boiler does not necessarily indicate that there is no emission problem. The coarse particles will fall out relatively close to the stack and on an equal weight basis will cause a greater local nuisance than fly ash from pulverized coal. Thus fine particles make a small plume appear much denser than coarse particles. It is for this reason that removal of particles larger than 20 microns has little effect on the appearance of the plume.

T A B L E 5-1

TYPICAL PARTICLE SIZE DISTRIBUTION OF FLY ASH FROM BOILER GASES OF VARIOUS BOILERS

(Percent by Weight)

Particle Size Microns	COAL FIRED BOILER					
	Pulverized Coal	Cyclone Furnace	Spreader Stoker	Traveling Grate	Underfed Stoker	100 Percent Bark Fired Boiler
0-10	25%	72%	11%		7%	12%
10-20	24	15	12		8	10
20-30	16	6	9	11	6	7
30-40	14	2	10		9	6
40-74	13		12	12	8	14
74-149	6	5	17	30	19	16
+149	2		29	47	43	35
TOTAL	100	100	100	100	100	100

Data from Bituminous Coal Research Association, Pittsburgh, Penn. and the Industrial Gas Cleaning Institute, "Criteria for the Application of Dust Collectors to Coal Fired Boilers."

Electrostatic Precipitators

The performance of an electrostatic precipitator is directly related to the level of power input maintained. The amount of electric charge that the dust particles will acquire depends upon the conductivity or resistivity of the dust. When the resistivity exceeds approximately 2 x 10^{10} Ohm centimeters excessive sparking is to be expected with subsequent drop in collecting efficiency. Research has shown that the resistivity of the fly ash is closely related to the presence of water, soluble sulfates, and free sulphuric acid formed on the surfaces of the dust particles by absorption of SO_3 from the flue gas. If the sulfur content is greater than 1 percent in the coal there is likely to be sufficient SO_3 in the flue gas to create favorable conditions for an acceptable resistivity. However, SO_3 is formed by oxidation of SO_2 which occurs between 750° and 1500° F. The amount of SO_3 formed will depend on the design of the boiler and the rate of combustion. The lower limit of soluble sulfur compounds in the coal ash, necessary to create an acceptable resistivity has been found to be between 0.5 and 1.0 percent. The resistivity also varies with the flue gas temperature and reaches a peak at approximately 300°F for average fly ash. The peak resistivity of some fly ash is high enough to adversely effect the precipitator performance. Yet 300°F is a favorable temperature level for boiler and evaporator operation. Thus, requirements for the boiler and for the precipitators are contradictory. The amount of magnesium and aluminum in the coal ash is also believed to be a factor contributing to high resistivity. Fly ash from strip mined coal usually has higher resistivity than fly ash from deep mined coal.

Fly ash containing more than 15 percent carbon will effect precipitator performances adversely. The reason for this is that carbon is a relatively good conductor and will acquire, but also loose, an electrical charge very easily. When a particle with high carbon content precipitates on the collector plate it loses its charge immediately. Owing to a phenomenon referred to as "charge by induction," the particle assumes a weak positive charge. It is repelled from the collector plate and reentrains into a gas stream. This phenomenon is repeated over and over again until the particle is carried out by the gas stream and escapes from the precipitator. In order to overcome this problem, the collector plates are given a special form and the so-called pocket or screen grid collecting electrodes have proven to be very effective. There is also a certain correlation between the carbon content and the coarseness of the fly ash. A coarse fly ash has normally a high carbon content.

Recent improvements in the design of electrostatic precipitators has reduced many of the advantages of combinations of electrostatic precipitators and mechanical collection systems, so common some years ago. The plate electrode precipitators did not have sufficient collecting efficiency on high carbon fly ash with problems of "snow outs" from the stack during soot blowing and collector plate rapping periods. Pocket electrodes and continuous rapping have, to a large extent, circumvented these problems. Another advantage of a combined system installation was that the mechanical collector would still remove a major portion of the fly ash should the precipitator be out of order for some reason. Sectionalized design and more reliable rectifiers have shortened the shut-down time considerably on modern precipitators.

Mechanical Collectors

Bark char is a difficult material to collect. While the particle size is relatively large, the specific gravity is very low and the sliver shape of the particles makes collection difficult. Bark char has a specific gravity in the 0.2 range; pulverized coal ash for example is 10-15 times heavier. In addition to the unique particle shape, the low specific gravity and the very fragile particles, the bark boiler collector must also be able to handle abrasive sand accompanying the bark. There are consequently a number of considerations to make when selecting a collector for a bark boiler.

Electrostatic precipitators are not generally suitable for use on bark boilers due to (1) the poor electrical characteristics of bark char and (2) the possibility of fires.

Some studies report that the bark char disintegrates into fine particles in a multi-tube collector due to the high centrifugal forces imposed. These fine particles coupled with their low specific gravity have a tendency to float through the outlet tube of the collector and out the stack. This phenomenon is also attributed to the short distance from the wall of the collecting tube to the outlet tube. A survey shows that there are more large diameter cyclone installations than multi-tube units operating on bark boilers in the wood pulping industry. It should be noted, however, that both collection devices can give sufficient efficiency providing they are selected properly.

Many bark char collectors in the southern part of the United States are operating with reinjection systems that are not equipped with provisions for separation of sand from the char before reinfection. Sand separation is essential for reducing excessive erosion of the collector and the boiler tubes. It is not unusual for a system operating on bark from southern pine to accumulate a number of truck loads of sand in an operating day. Some mills are now selling the collected bark char for manufacturing of charcoal briquettes.

In many multi-tube collectors recovery vanes are installed to reduce pressure drop and collector size. The experience shows that collectors operating on boilers firing oil in combination with bark must be designed without recovery vanes. Some oil fractions having high dewpoints combine with the bark char particles and plug the recovery vanes. One single plugged outlet tube can by-pass some 10 times its normal share of flue gas uncleaned into the stack. It is consequently imperative to prevent plugging of the tubes.

Studies show that substantial improvement of the collecting efficiency of bark char collectors can be attained by changing the normally conical bottom outlet of the cyclone (tube). By using straight cylindrical tubes with a peripheral discharge rather than conventional conical discharge, the ash is removed from the tube before the gas reverses into the inner vortex and in sufficient distances from the turning point. This prevents re-entrainment of already collected particles.

The volume handled by a given diameter collector tube depends on the shape of the inlet guide vane. Shallow pitched vanes give more spin and handle less volume. Steep pitched vanes reduce the spin and can handle more volume. The shape of the inlet guide vane can make a difference in volume capacity of the tube at a ratio of 1 to 2. The vanes are of obvious importance for collector size as well as operating efficiency.

Bark char collectors should be easily accessible for inspection, maintenance and repair because of the difficult operating condition under which they work. One always has to bear in mind the abrasiveness of the dust and possibility of clogging. Large diameter cyclone collectors are easily accessible for maintenance. Some modern multi-tube collectors also offer design with accent on maintenance. The place subject to wear in a well designed collector is limited to the collection tube or cyclone cone. Hard cast iron is often used in tubes. Hardness of 420 Brinell has proven to be a maximum, because hard tubes are subject to thermal shock damage. Cyclone cones, as, well as the entire cyclone, is sometimes furnished with abrasion resistant linings. This lining is usually installed over a hexagonal steel liner.

Both mechanical collectors and electrostatic precipitators are sensitive to uneven gas and dust distribution. The entrance to the collector is normally provided with devices for correction of uneven gas distribution. The space for proper distribution ahead of the collector is almost always very limited. Sharp bends and short transitions tend to produce uneven distribution across the entrance. Various devices such as spreaders, turning vanes and perforated baskets in the collector entrance are used in efforts to overcome these problems.

Hopper fires occur in bark char collectors. Unlike ash from coal fire boilers, bark char must be continuously and completely removed. A small air leakage is enough for accumulated char to catch fire. Gas tight construction is essential in bark char collectors.

Design must consider the possibility and effect of fires. This removal from the hopper must be accomplished without a back flow of air. This can be done must successfully with rotary steel valves.

Bark boiler dust collectors should be carefully sized for correct pressure drop inlet velocity. While efficiency generally increases with increased pressure drop (and velocity) there are limits beyond which fragile material like bark char will be broken into smaller particles which has a detrimental effect on the collecting efficiency.

The low specific gravity of bark char, when the boiler is operated on 100 percent bark without reinjection, or the finer particle size distribution when the ash is reinjected, create operating conditions which limit the application of large diameter cyclones on these installations.

Only when other conditions are superimposed over these conditions can the large diameter cyclone approach 92 percent efficiency in a single stage unit. One of the conditions which will increase the efficiency is the amount of sand in the bark char.

When operating on 100 percent bark, the efficiency will vary greatly depending upon the amount of sand imbedded in the bark. Without reinjection, much of the imbedded sand remains in the bark char. This greatly increases the average specific gravity of the bark char, making the centrifugal separation easier with higher efficiencies.

Example: 80% Bark	Specific Gravity (as char)	0.3
20% Sand	Specific Gravity	2.8

$$S.\ G. = (0.8)\ (0.3) + (0.2)\ (2.8) = 0.24+0.56 = 0.80$$

For a large diameter cyclone where the specific gravity is assumed at 0.3, the most practical unit would give efficiencies in the 70 to 75 percent range. If, as in the example above, the char contained 20 percent sand, the efficiency would increase to 85 to 89 percent.

The performance of mechanical collectors, with 40 percent bark and 60 percent oil feed to the boiler, is greatly improved due to the higher specific gravity of the flyash. The nature of this dust, however, in multi-tube collectors creates a plugging problem. The higher specific gravity and tendency not to plug make the large diameter cyclone practical for combination fired boilers. It is possible to achieve the 92 percent design efficiency in both multi-tube and large diameter cyclones with single-stage units. The 96 percent design efficiency is possible with two-stage large diameter or multi-tube units.

5.3.6.7 Combination Boilers

Application

The following methods are considered for reducing emissions from an existing combination boiler:

1. Replace existing multi-tube collector with new design multi-tube collector(s).
2. Add a cyclonic scrubber in series with an existing 85 percent dust collector.
3. Addition of surge bin to reduce peak bark flows from the woodyard.
4. Convert 100 percent reinjection system to 70 percent or less.

The application of equipment for the first two methods has been discussed in the preceeding section.

Addition of Surge Bin to Reduce Peak Bark Flows. Particulate emissions from a combination boiler burning bark and oil are approximately proportional to the rate of bark feed to the boiler, assuming other variables are approximately constant. In order to minimize particulate emission and insure good combustion conditions to the furnace, current practice is to install a surge bin with a new combination boiler. Many existing mills feed bark directly from the woodyard to the boiler without a surge bin. For a number of reasons, the bark flow rate from a woodyard fluctuates widely, and bark supply does not always coincide with fuel demands.

The installation of a properly sized and designed surge bin will provide a relatively constant bark flow to the combination boiler. The surge bin size may vary considerably for a particular boiler depending upon the relative woodyard debarking capacity in relation to the particular boiler's bark burning capacity.

While surge bins are usually installed to obtain more efficient use of the combination boiler and better combustion conditions, a definite reduction in particulate emission will also result. Where particulate emissions must be reduced from a combination boiler, the installation of a "surge bin" should be evaluated.

Convert 100 Percent Reinjection System to Approximately 70 Percent or Less. Combination boilers burning bark and oil may utilize bark char reinjection from the boiler hoppers (located below the mud drum), dust collectors hoppers, and/or ductwork hoppers. The bark char results from the incomplete combustion of the bark fuel. Reinjection is practiced primarily to eliminate disposal problems with bark char and to obtain the additional heat from burning the char.

Current practice in the design of combination boilers is to use reinjection systems which reinject 70 percent or less of the collected bark char. A few older mills still practice 100 percent reinjection. An assessment of the impact of reinjection has been investigated in detail by Mullen (11). In essence, 100 percent reinjection of ash particles results in a continuous cycling of the ash. This continual cycling degrades the ash particle size until the ash is emitted from the stack, thereby increasing particulate emission. For bark which contains extremely low ash (less than 1.0 percent) the influence of 100 percent reinjection may be minor. This is especially true where the bark and bark char are burned in refractory furnaces.

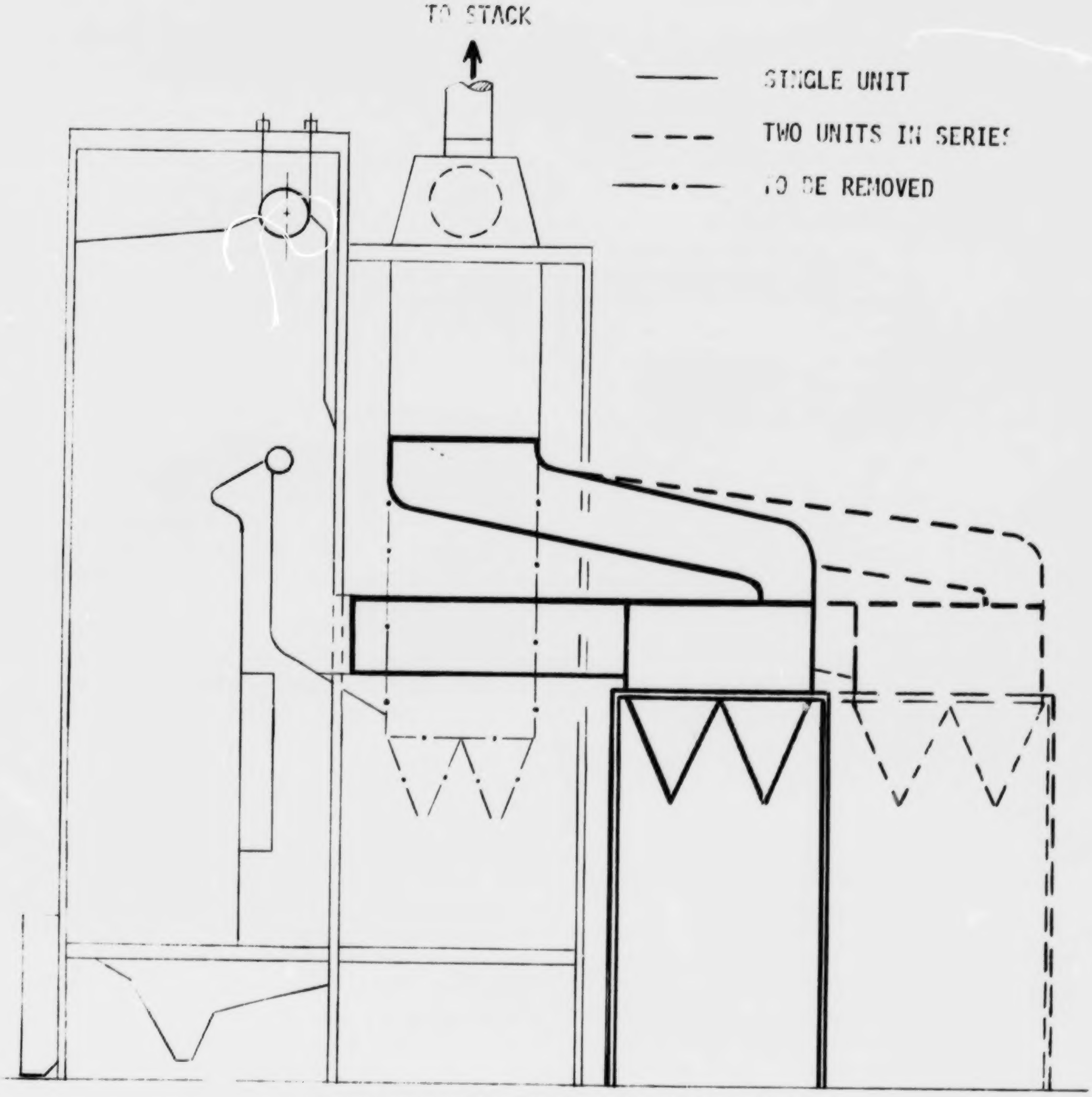
Where particulate emission from a combination boiler is high and 100 percent bark char reinjection is practiced, reduction of reinjection to 70 percent or less should be evaluated. The cost of reducing the percentage reinjection will vary considerably from mill to mill, and should be developed on an individual mill basis. The cost will be expected to be very much higher where an ash disposal system must be installed.

Costs

New Design Multi-tube Cyclones. This method is based on the replacement of an existing relatively low efficiency, multi-tube cyclone collector with a new design, high efficiency, multi-tube cyclone system, two cases being considered. The first case assumes the replacement of the existing system with a single stage system, and the second case assumes the replacement of the existing system with a two-stage system. Both are shown in Figure 5-66.

It was assumed that 40 percent of the heat input to the boiler was from bark and 60 percent from oil.

Capital costs are based on mild steel construction for the collector and duct work. Erection costs are based on components being shipped knocked-down. It was also assumed that existing sand classifiers could be used on the new single stage unit and that classifiers would not be required on the secondary stage collector. The ash collected in the second stage would be discharged directly to the existing ash handling system.


TO STACK
SINGLE UNIT
TWO UNITS IN SERIES
TO BE REMOVED

ADDED EQUIPMENT SHOWN WITH HEAVY LINES

FIG. 5-66 TYPICAL ARRANGEMENT FOR REPLACING AN EXISTING SINGLE STAGE MULTI-TUBE COLLECTOR WITH A NEW SINGLE STAGE OR TWO STAGE MULTI-TUBE COLLECTOR - COMBINATION BOILER

Capital costs and net annual cost are presented in Figures 5-67 and 5-68. For a specific combination boiler, it should be recognized that arrangement and location of replacement collector system would vary widely, but the prepared estimates are typical of the costs that could be encountered.

Cyclonic Scrubber. This control method is based on the installation of one or more cyclonic scrubbers depending on combination boiler exit gas volume, with the scrubbers in series with and downstream of the existing dust collector as shown in Figure 5-68.

In addition to the scrubber, a circulation pump and motor for each scrubber, a strainer for removal of ash and char from cyclone effluent, required piping for make-up, circulation, and effluents to ash lagoon and waste disposal system, reivsed ductwork to connect the scrubber with the existing system, and instruments and controls for proper operation of the above are included.

The ash and char may be collected in containers or discharged into the existing ash handling system. For the purposes of this method, the ash and char is assumed to be discharged to the existing ash handling system.

The capital cost is based on mild steel construction with concrete lining for the cyclonic scrubber and ductwork as this would be the most representative material for this service. Cost curves are shown in Figure 5-70.

Effectiveness

Particulate Removal. Efficiencies for particulate removal are predicted as follows:

Control Method	Annual Operating Efficiency
Single-stage Collector	90.0%
Two-stage Collector	94.0%
Cyclonic Scrubber	94.0%

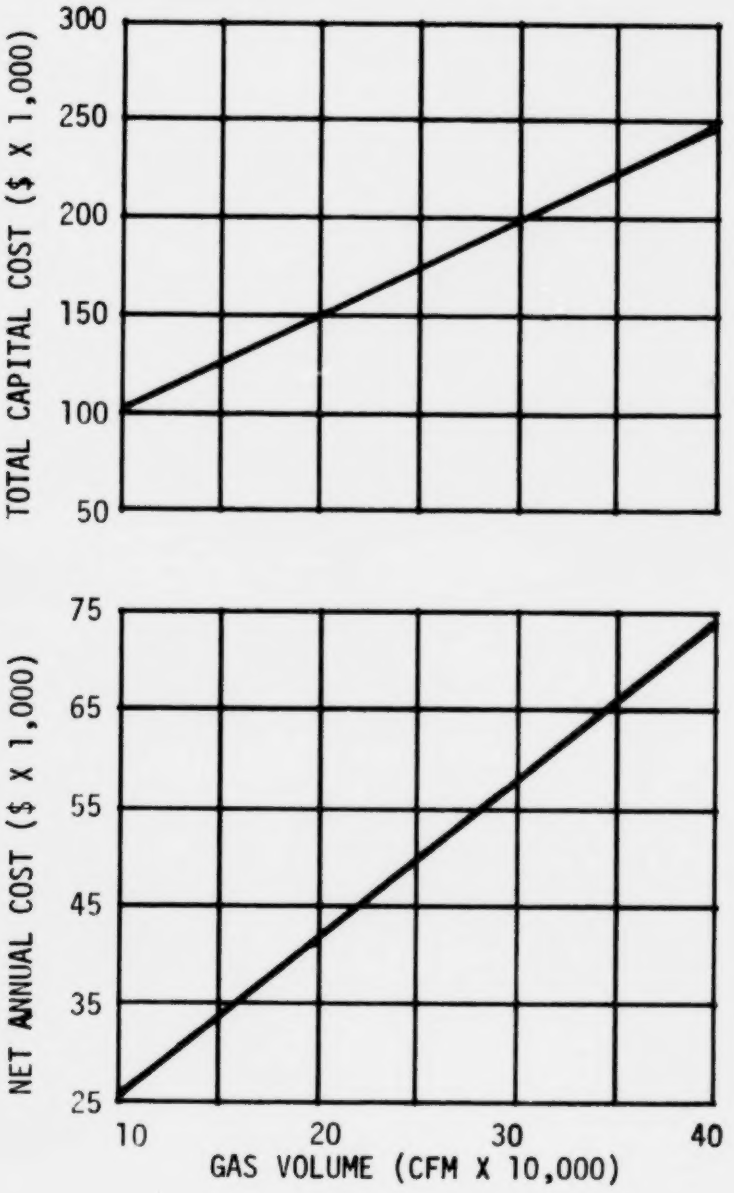


FIG. 5-67 CONTROL METHOD COSTS FOR REPLACING AN EXISTING MULTI-TUBE COLLECTOR WITH A SINGLE-STAGE COLLECTOR-COMBINATION BOILER (90.0% A.O.E.)

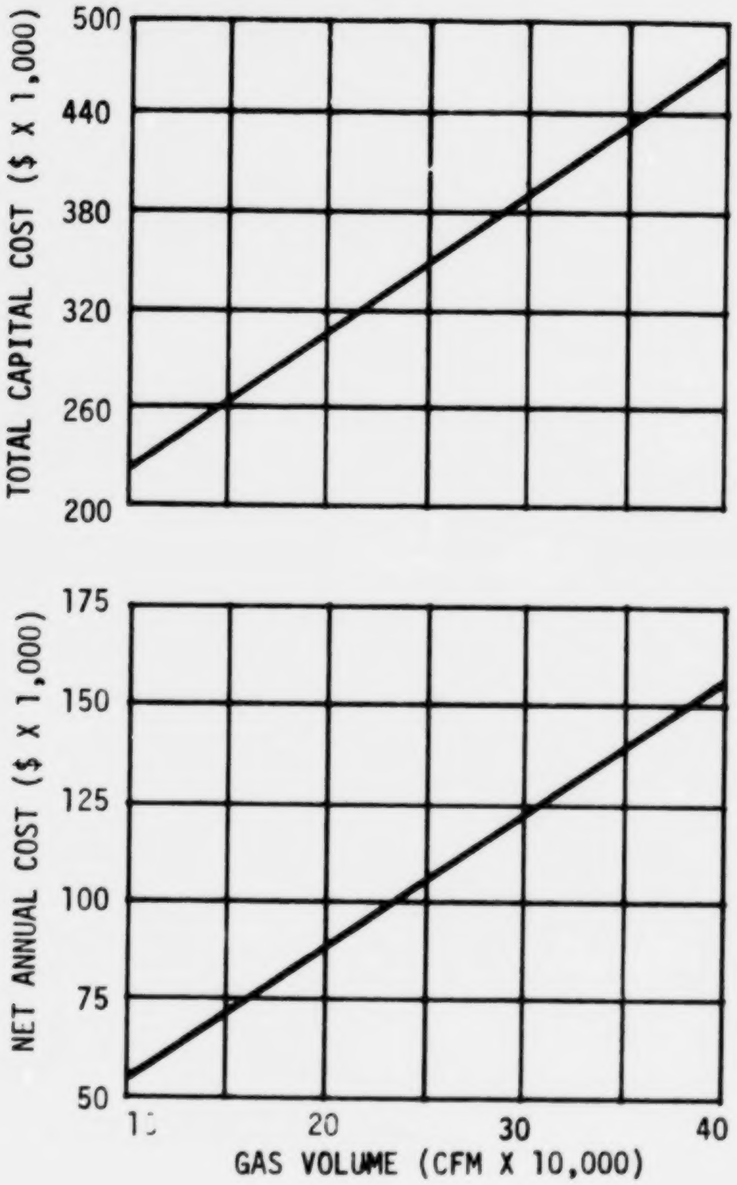
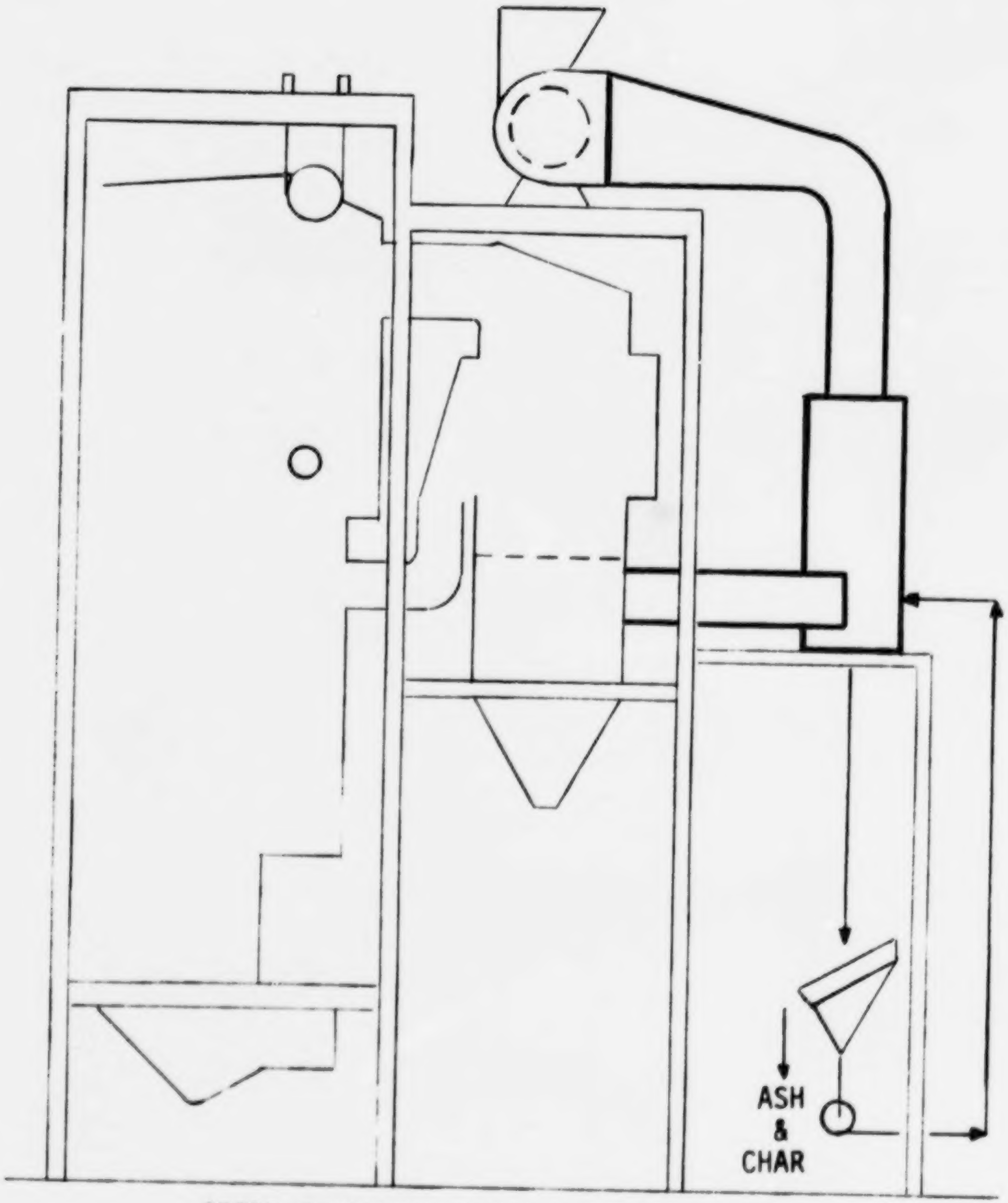


FIG. 5-68 CONTROL METHOD COST FOR REPLACING AN EXISTING MULTI-TUBE COLLECTOR WITH A TWO-STAGE COLLECTOR - COMBINATION BOILER (94.0% A.O.E.)



ADDED EQUIPMENT SHOWN WITH HEAVY LINES

FIG. 5-69 TYPICAL ARRANGEMENT FOR CYCLONIC SCRUBBER ADDED IN SERIES WITH AN EXISTING 85% EFFICIENT DUST COLLECTOR - COMBINATION BOILER

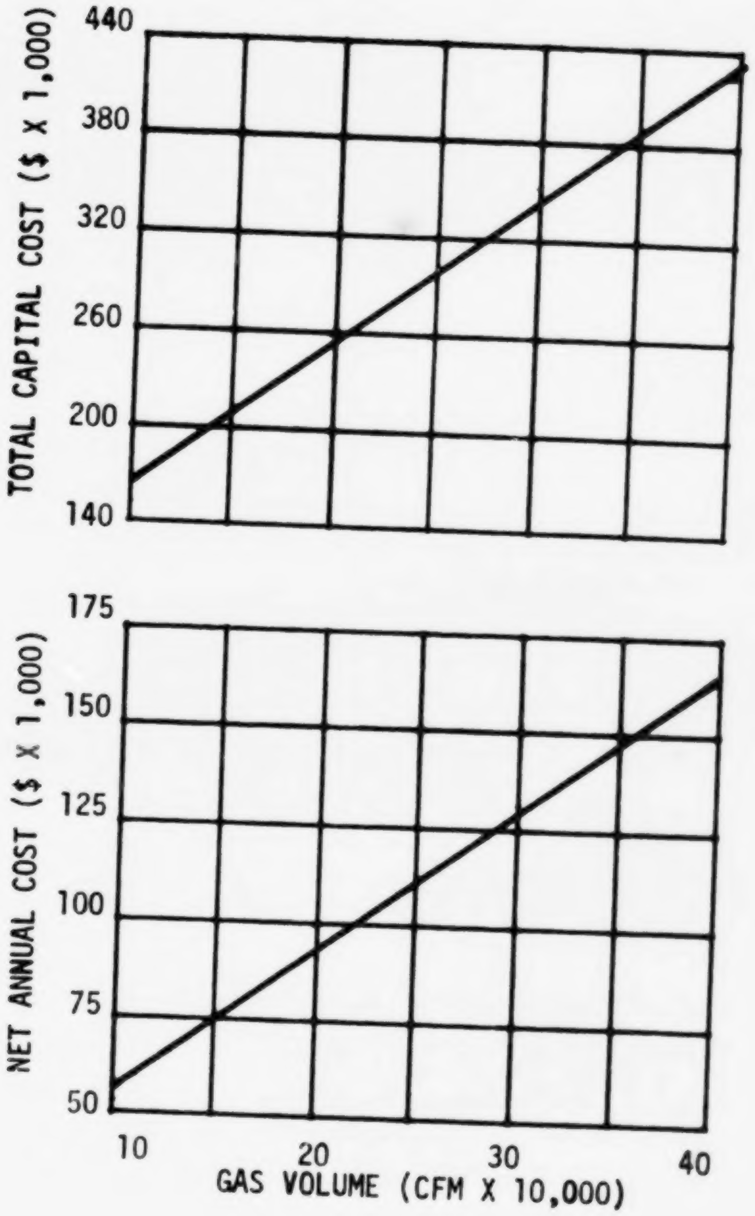


FIG. 5-70 CONTROL METHOD COSTS FOR A CYCLONIC SCRUBBER ADDED TO AN EXISTING 85% EFFICIENCY DUST COLLECTOR - COMBINATION BOILER (94.0% A.O.E.)

Particulate removal efficiencies for the remaining methods are not predictable.

Curves for particulate cost effectiveness are presented in Figures 5-71 and 5-72.

Reduced Sulfur and Sulfur Dioxide Removal. None of the methods considered would be effective in reducing emission of sulfur dioxide.

Operation. In comparing operation the cyclonic scrubber would result in more corrosion potential and might also be more prone to plugging. In considering surge bins, arching of the bark can be an operating problem.

Summary. Due to the range in efficiencies and costs, the selection of the most effective method would have to be determined for each individual case.

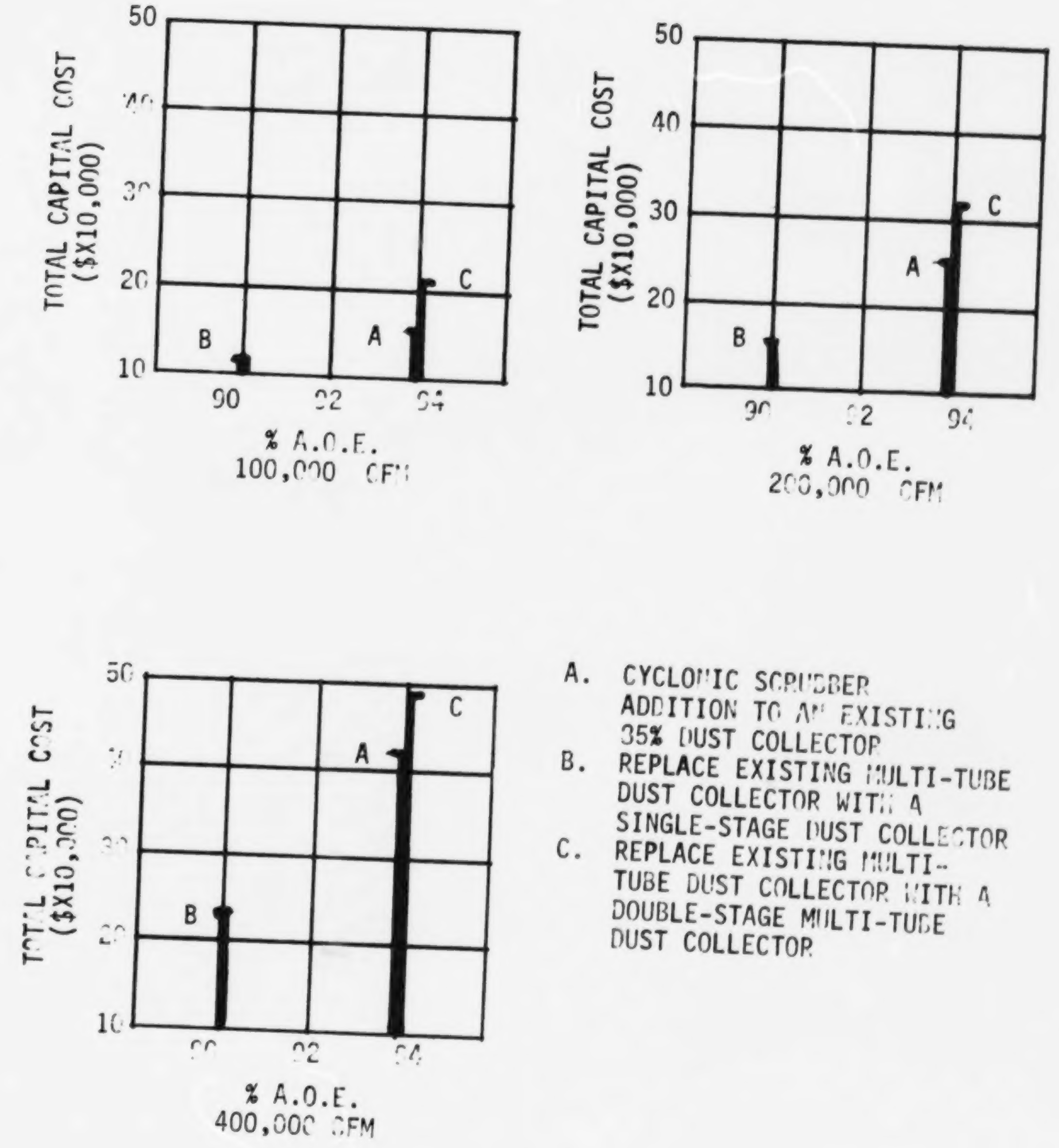


FIG. 5-71 PARTICULATE COST EFFECTIVENESS (TOTAL CAPITAL COST COMPARISON) FOR CONTROL EQUIPMENT ADDED TO COMBINATION BOILER

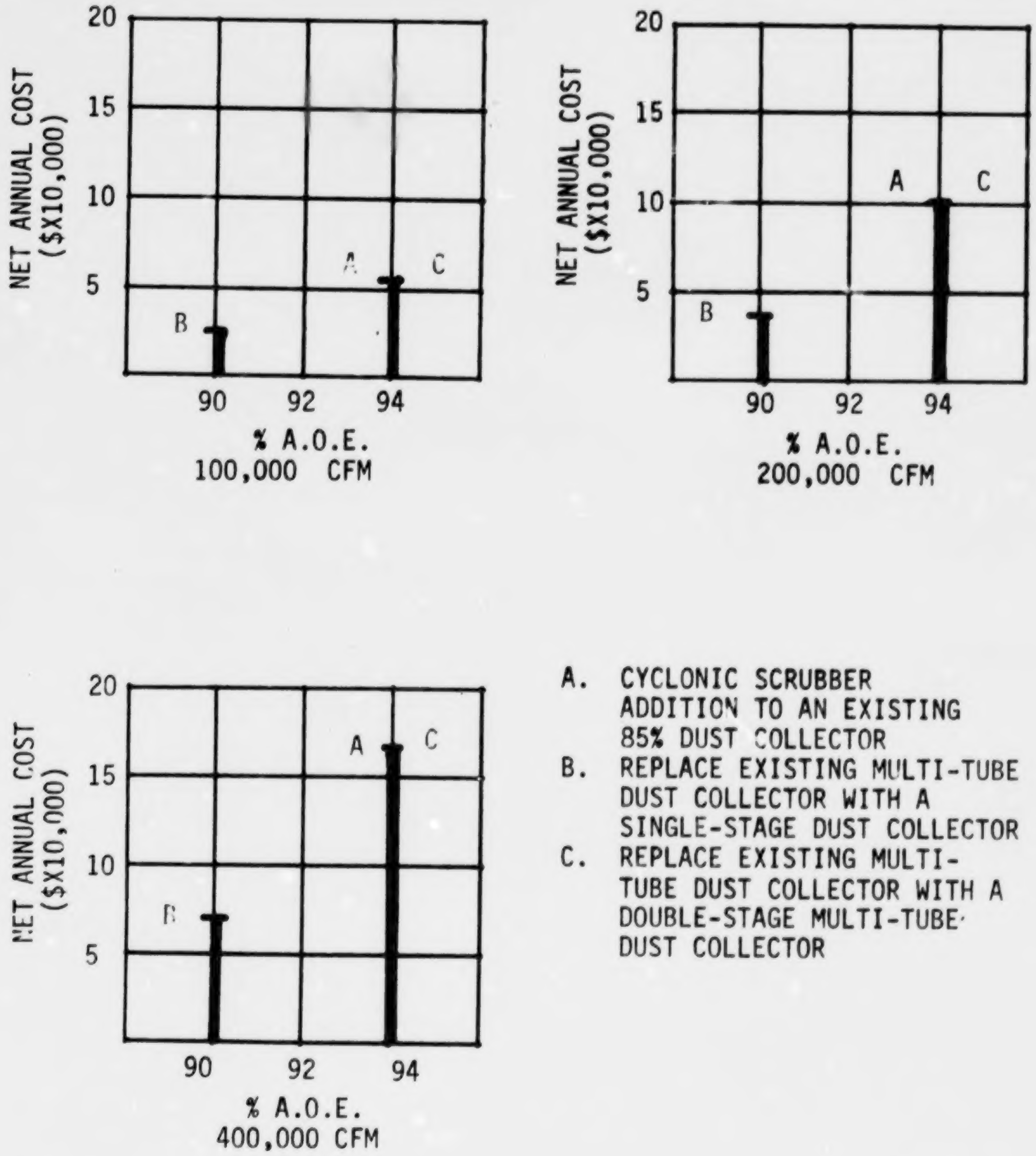


FIG. 5-72 PARTICULATE COST EFFECTIVENESS (NET ANNUAL COST COMPARISON) FOR CONTROL EQUIPMENT ADDED TO COMBINATION BOILER

5.3.6.8 Power Boilers

Application

This case is based on the addition of an electrostatic precipitator following an existing dust collector for a coal-fired power boiler. Two precipitator sizes are considered.

Costs

Estimated costs are based on the equipment arrangement shown in Figure 5-73. Curves for capital costs and net annual operating costs are presented in Figures 5-74 and 5-75.

The equipment costs are based on a double chamber, steel shell precipitator. All labor and material for a complete installation are included in the capital cost and the net annual operating cost. The precipitator size is based on firing coal containing 2 percent sulfur.

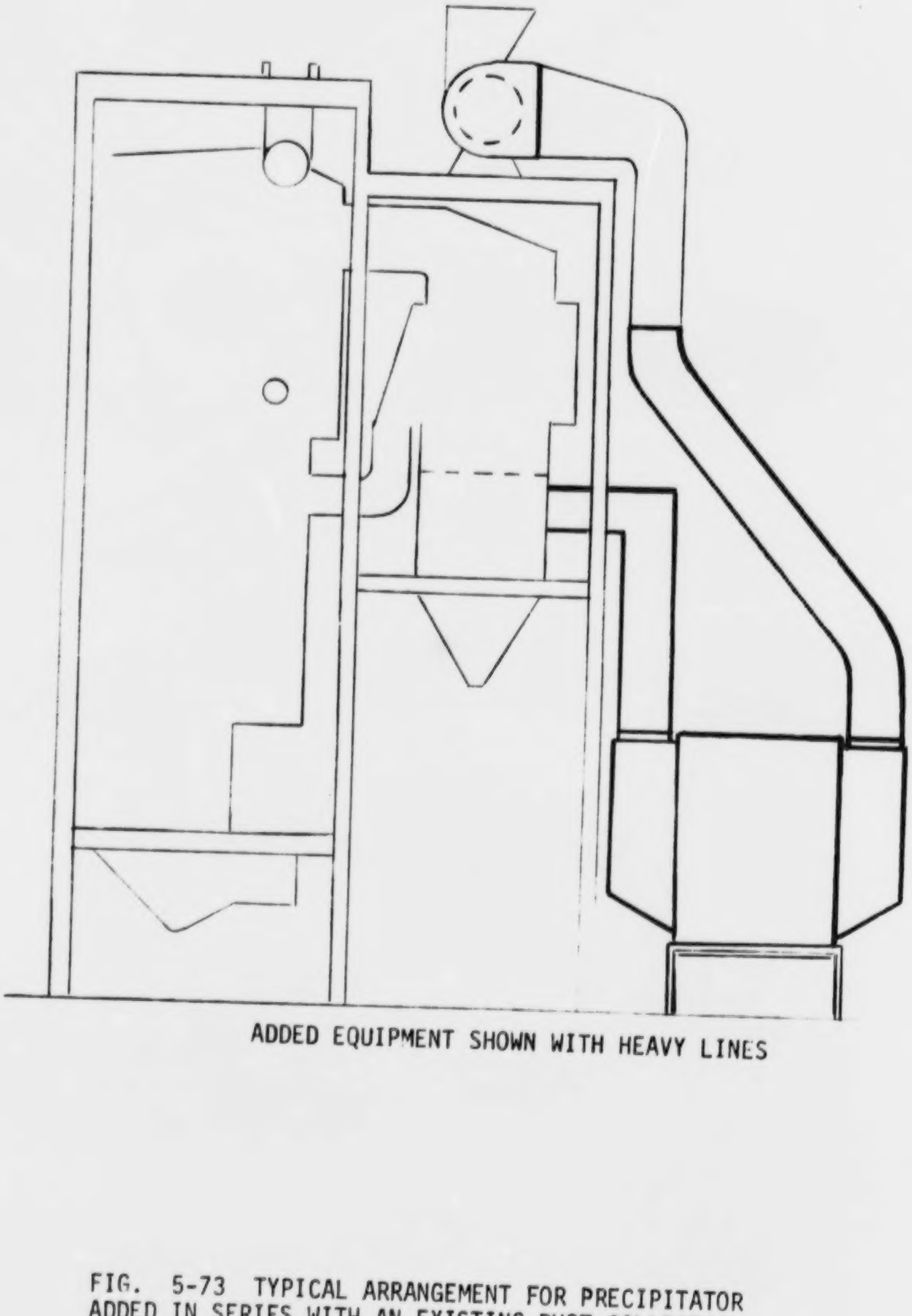
Effectiveness

Particulate Removal. Guaranteed particulate efficiencies for the two precipitators considered are 99.0 percent and 90.0 percent. The resulting total annual operating efficiencies are:

Precipitator Guaranteed Efficiency	Precipitator AOE	Dust Collector AOE	Total AOE
99.0%	98.0%	80.0%	99.0%
90.0%	89.0%	80.0%	98.0%

Reduced Sulfur and Sulfur Dioxide Removal. Precipitators are ineffective for removing sulfur compounds.

Operation. The operating characteristics of the precipitators should be identical.



ADDED EQUIPMENT SHOWN WITH HEAVY LINES

FIG. 5-73 TYPICAL ARRANGEMENT FOR PRECIPITATOR ADDED IN SERIES WITH AN EXISTING DUST COLLECTOR - COAL FIRED POWER BOILER

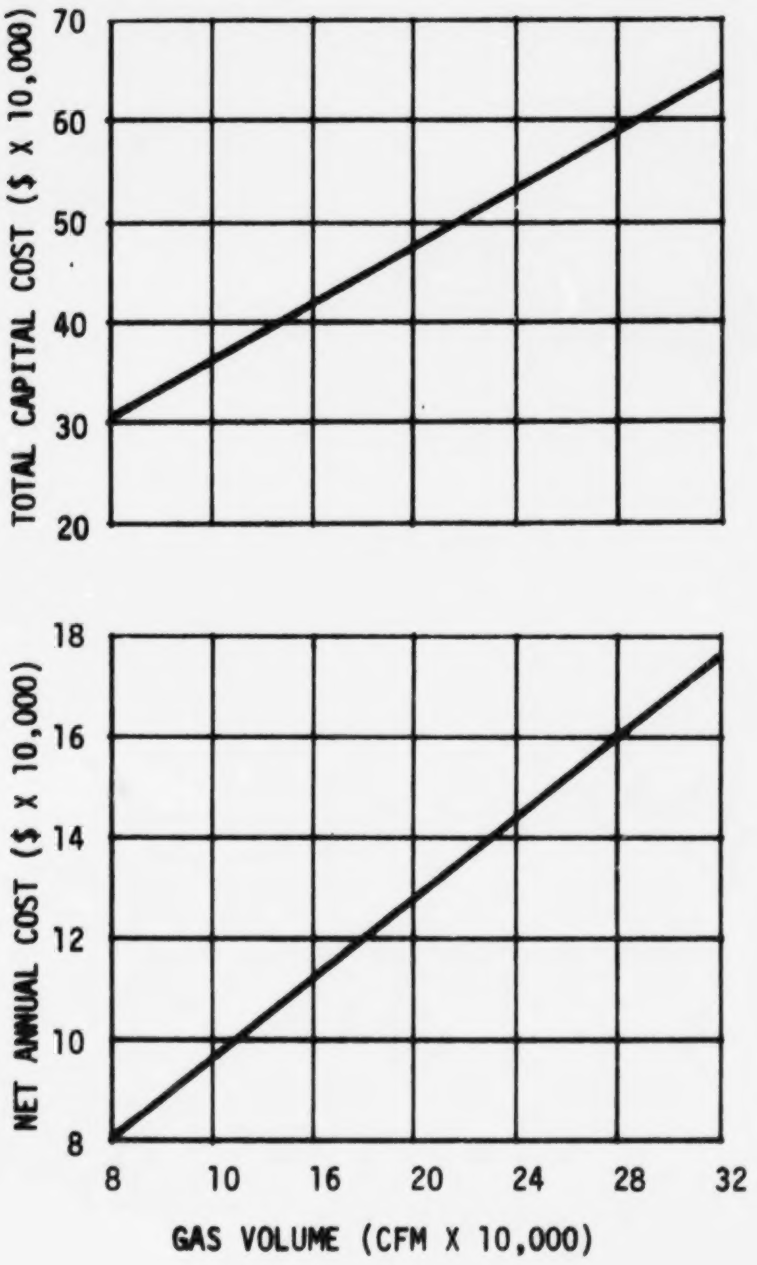


FIG. 5-74 CONTROL METHOD COSTS FOR 99.0% EFF. ELECTROSTATIC PRECIPITATOR ADDED TO AN EXISTING 90% EFF. DUST COLLECTOR COAL FIRED POWER BOILER (99.0% A.O.E.)

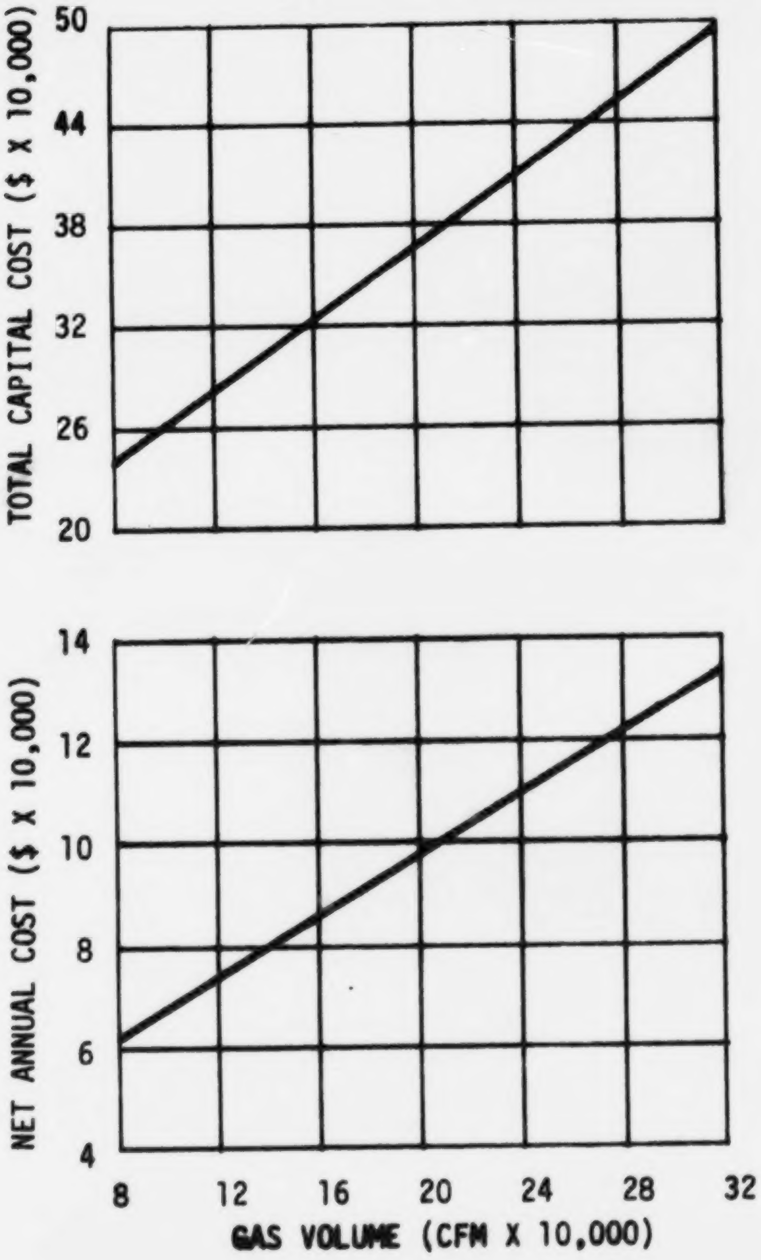


FIG. 5-75 CONTROL METHOD COSTS FOR 90.0% EFF. ELECTROSTATIC PRECIPITATOR ADDED TO AN EXISTING 90% EFF. DUST COLLECTOR COAL FIRED POWER BOILER (98.0% A.O.E.)

5.3.7 SULFITE SOURCES

The evaluation of control methods for sulfite sources is limited to those which are distinct from the kraft sources. The reader is referred to the previous discussions of kraft sources for the following sulfite sources: Washer vents, evaporators, power boilers, and combination boilers. It should also be noted that control methods have not been considered for the following sulfite sources: Dump tank vent, Venturi absorbers, absorption towers, and ammonia incineration. These sources are either of minor importance from an emission standpoint, or there are no control methods presently in use.

5.3.7.1 Acid Tower

Application

The pressure accumulator is vented into the acid storage tank because the two tanks are nearby. The acid storage tank is then vented to the absorption tower. This system effectively prevents emissions from these items of equipment, thereby leaving the acid absorption tower as the only significant source of emission in the acid system.

The construction of most absorption towers incorporates a mesh pad distribution system for absorption liquid. Therefore, additional mesh pads are deemed necessary.

The efficiency of absorption of most sulfiting towers exceeds 90 percent. Some mills have placed a second absorption tower in series with the sulfiting unit for scrubbing exhaust gases. This method has been tried only where an existing second tower was available and could be used economically. An example would be the conversion of calcium base, requiring two towers, to ammonium base which requires one tower.

A cost estimate has been prepared for installation of a second packed tower in series with the absorption towers.

Costs

The capital costs are based on purchase and installation of a new absorption tower in series with an existing acid absorption tower. Costs are predicated on installation of the new tower in an existing mill with a minimum of down time for placing the new tower on stream. Capital and net annual costs are shown in Figure 5-76.

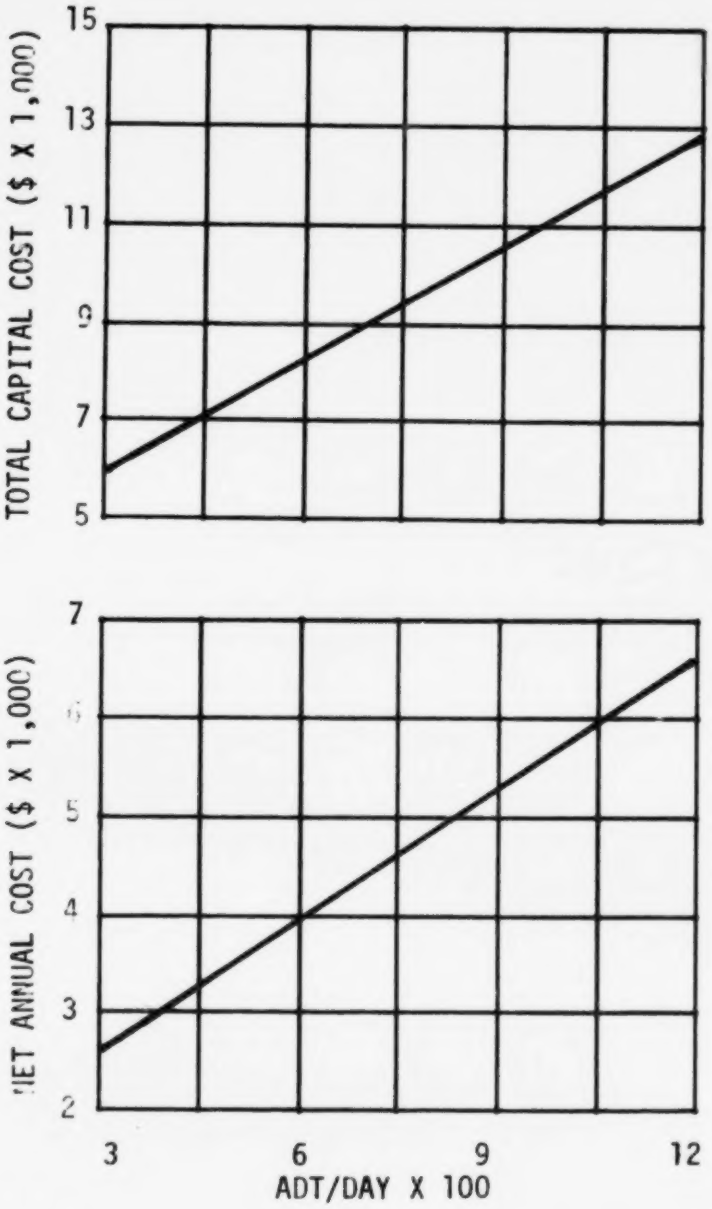


FIG. 5-76 CONTROL METHOD COSTS FOR PACKED TOWER ADDED TO ACID TOWER

Effectiveness

Data are not presently available on economic feasibility or efficiency of recovery of this system.

5.3.7.2 Blow Pit

Application

Two systems were considered to replace the existing multiple wooden blow stacks and showers with a high efficiency SO_2 recovery system as follows:

1. Condenser with Cyclone and Absorption Tower
2. Packed Tower

Flash steam, sulfur dioxide, and inert gases are released in the blow pit during a digester blow. These gases then exit from the blow pit and after the steam is condensed, the noncondensible gases, mainly sulfur dioxide, are absorbed in a packed tower. The recovered sulfur dioxide is reused in the process and the condensate creates a source of hot water.

Costs

The costs of these controls are comprised chiefly of the purchase and installation of equipment for condensing the flash steam and a packed tower for absorbing the sulfur dioxide. In addition, the costs include piping for conveying the gases from one stage to another and carrying the water required for gas absorption and heat transfer. The costs for these control methods are "Savings" not costs and are shown in Figures 5-77 and 5-78.

Effectiveness

Particulate Removal. There are no particulate emissions from the above source.

Reduced Sulfur and Sulfur Dioxide Removal. There are no reduced sulfur emissions from the above source. Both systems are 95 percent efficient in absorbing sulfur dioxide from the blow pit stacks, where the SO_2 concentration is approximately 4 percent during the initial stage of the blow. This produces recovery water containing 0.85 percent sulfur dioxide. The recovery water is subsequently used directly in the fortifying of the cooking acid or the SO_2 is stripped from the recovery water and used in the manufacture of cooking acid.

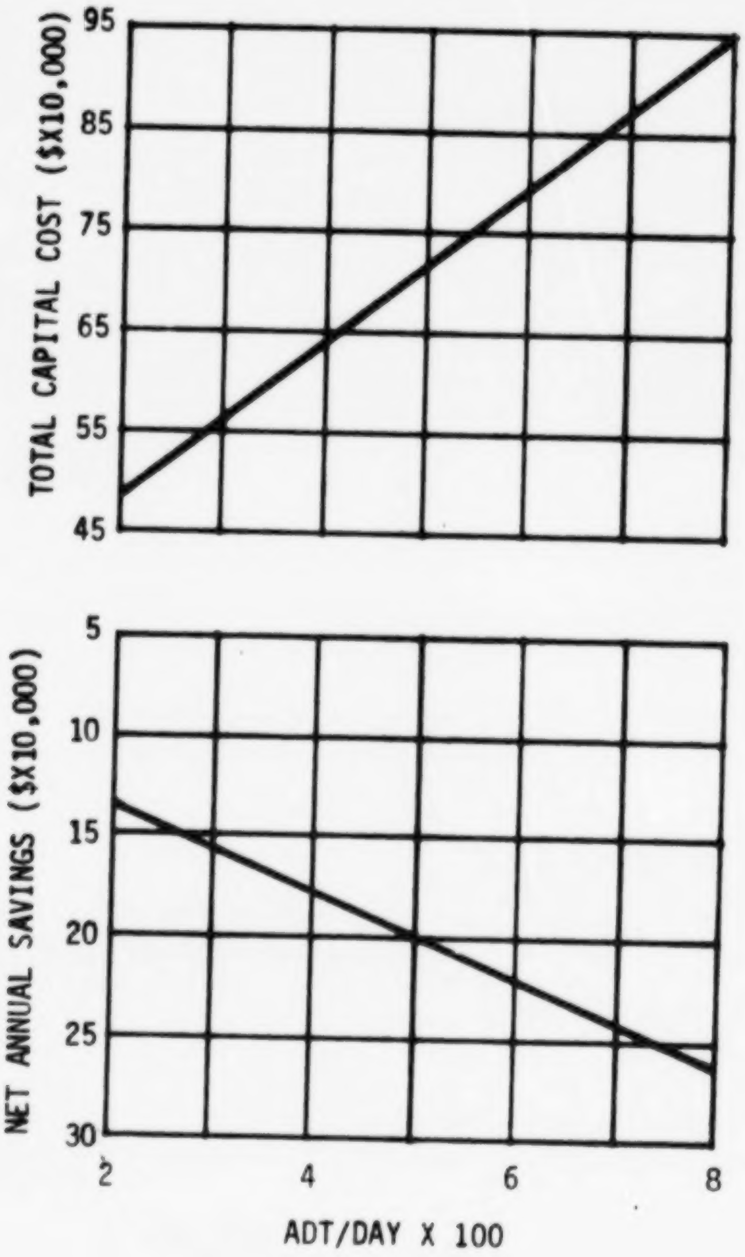


FIG.5-77 CONTROL METHOD COSTS FOR BLOW PIT - CONDENSER WITH CYCLONE AND PACKED TOWER

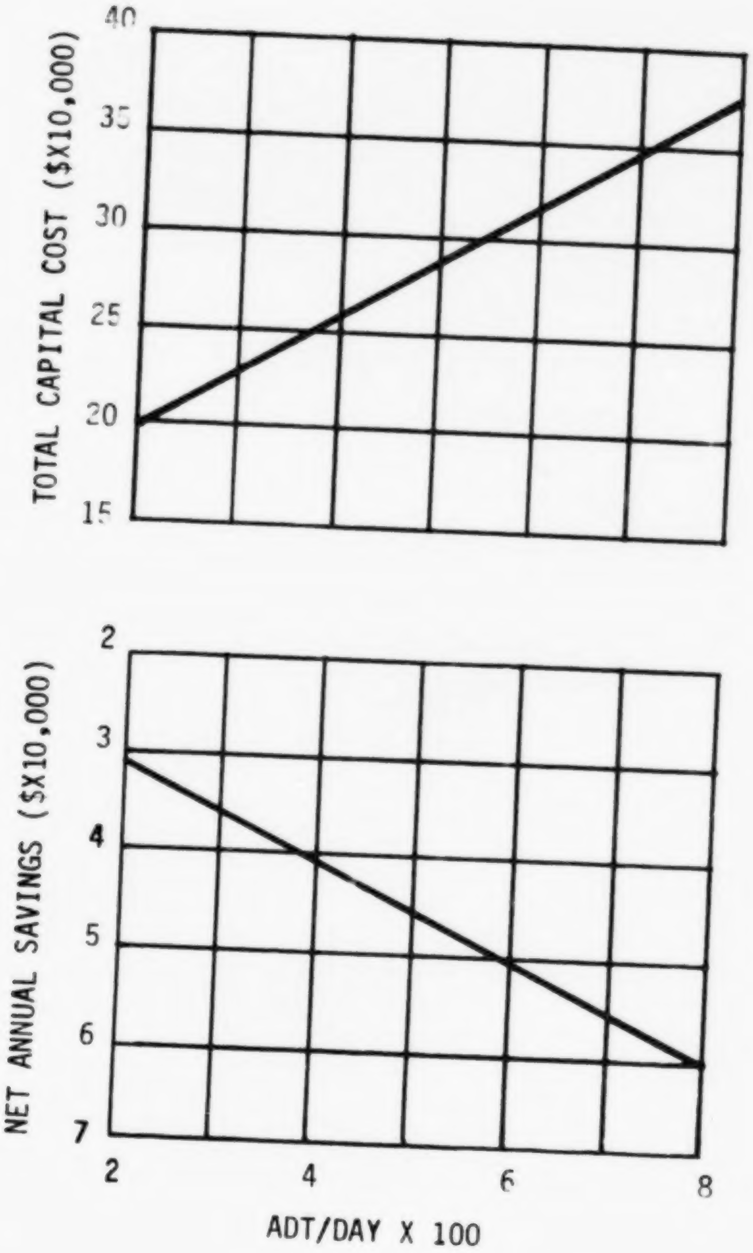


FIG.5-78 CONTROL METHOD COSTS FOR BLOW PIT - PACKED TOWER

Operation. System (a) may have more problems than (b) because of additional equipment and instrumentation. System (b) consists mainly of a gas header and an absorption tower, and is, therefore, relatively trouble free in operation.

Summary. The capital cost of System (a) is approximately three times the cost of System (b), both having a sulfur dioxide recovery efficiency of 95 percent. Although System (a) has a higher heat recovery efficiency, it would appear that System (b), Packed Tower, is more effective.

5.3.8 NSSC SOURCES

Evaluations of control methods applied to NSSC sources are not included due to the lack of emission data and application experience. The reader is referred to the previous Kraft discussions for the following NSSC sources: Washer vents, evaporators, combination boiler, and power boiler.

11. Mullen, J. F., "A Method for Determining Combustible Loss, Dust Emission, and Recirculated Refuse for a Solid Fuel Burning System," Paper presented at ASME Winter Annual Meeting, New York, November 29 - December 4, 1964.

12. Wrist, Peter, Mead Corporation, Verbal Communication, Liaison Meeting, New York, November 6, 1969.

13. Carlton-Jones, Dennis and Schneider, H.B., "Tall Chimneys," Chemical Engineering, 75, October 14, 1968.

14. Martin, F., "Secondary Oxidation Overcomes Odor from Kraft Recovery," Pulp and Paper, 43, June 1969.

15. Industrial Gas Cleaning Institute, "Test Procedure for Gas Scrubbers," Publication No. 1, IGCI, Box 448, Rye, New York, 1964.

16. Industrial Gas Cleaning Institute, "Criteria for Performance Guarantee Determinations," Publication E-P3, IGCI, Box 448, Rye, New York, 1965.

17. NAPCA, "Tall Stacks - Various Atmospheric Phenomena and Related Aspects," Publication No. APTD 69-12.

18. Wrist, Peter, Mead Corporation, Personal Communication, Sept. 2, 1969.

19. Blosser, R. O., and Cooper, H.B.H., "Current Practices in Thermal Oxidation of Noncondensible Gases in the Kraft Industry," Atmospheric Pollution Technical Bulletin No. 34, National Council for Air and Stream Improvement, Inc., New York.

5.5 REFERENCES

1. Shah, I. S. and Stephenson, W.D., "Weak Black Liquor Oxidation System: Its Operation and Performance," TAPPI 51 (a), 87-, 1968.
2. "Steam, Its Generation and Use," 37th Edition, The Babcock and Wilcox Company, New York.
3. Fryling, G. R., "Combustion Engineering," Revised Edition First Impression, Combustion Engineering, Inc., New York.
4. MacDonald, R. G., Editor, "Pulp and Paper Manufacture - Volume I, The Pulping of Wood," Second Edition, McGraw-Hill, New York 1969
5. Whiteny, R. P., Editor, "Chemical Recovery in Alkaline Pulping Processes," TAPPI Monograph Series No. 32, TAPPI, New York, 1968
6. Theon, G. N., DeHaas, G. G., Tallent, R. G., and Davis, A. S., "The Effect of Combustion Variables on Release of Odorous Compounds from Kraft Recovery Furnaces," TAPPI 51 (8), 329-, 1968.
7. Harding, C. I. and Galeano, S. F., "Using Weak Black Liquor for Sulfur Dioxide Removal and Recovery," TAPPI 50 (10), 48A-, 1967.
8. Tomlinson, G. H., Chapter 5, page 419, "Pulp and Paper Manufacture - Volume I, Preparation and Treatment of Wood Pulp," First Edition, McGraw-Hill, New York, 1950.
9. Thomas, E., Broadus, S., and Ramsdell, E. W., "Air Pollution Abatement at S. D. Warren's Kraft Mill in Westbrook, Maine," TAPPI 50 (8), 81a-, 1967.
10. Hough, G. W. and Gross, L.J., "Air Emission Control in a Modern Pulp and Paper Mill," American Paper Industry, 36-, February 1969.

CHAPTER 6

NEW DEVELOPMENTS IN CONTROL TECHNOLOGY

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CHAPTER 6

NEW DEVELOPMENTS IN CONTROL TECHNOLOGY

SUMMARY

New control methods evaluated in this chapter are those which have had limited or no application in the United States, but which may show promise for more economically or effectively reducing atmospheric emissions. Only methods which promise improvement over existing technology have been evaluated in depth. Due to the indeterminate characteristics of some of the new methods, only order of magnitude costs have been estimated in order to make a relative comparison of the new techniques.

The following systems have been evaluated:

Kraft Recovery Furnace	- Conversion to High Solids System
	- Conversion to Air Contact Evaporator System
	- Sodium Carbonate - Bicarbonate Scrubber System
	- High Solids Plus Brine Scrubber System
	- Weak Black Liquor Oxidation (Molecular Oxygen)
Black Liquor Oxidation Unit	- Packed Tower Scrubber
	- Orifice Scrubber
Smelt Dissolving Tanks	- Incineration in Recovery Boiler
Digester Relief and Blow Plus ME Evaporators	- Separate Thermal Oxidation
Combination Boilers	- Mechanical Collector Plus Shave-off Scrubber
Brown Stock Washers	- Continuous Diffusion Washing
	- Enclosed Pressure Washing
	- Incineration
Ammonium Sulfite Liquor	- Incineration in Combination Boiler
	- Incineration in Separate Boiler

6.1 INTRODUCTION

In this chapter control methods which have had limited or no application within the United States, but which may show promise for economically and/or effectively reducing atmospheric emissions are investigated. Only methods which indicate possible improvement over existing technology will be evaluated in depth. All methods in this chapter have not been proven in U. S. practice and should be considered as experimental until their practical performance has been demonstrated under full-scale operating conditions.

Due to the indeterminant characteristics of new developments, capital cost and net annual cost curves have not been prepared. However, an assessment of capital cost and net annual cost have been made to obtain order of magnitude costs for a relative comparison of new developments. This order of magnitude cost is used to compare the relative control effectiveness of new developments with the present control methods described in Chapter 5.

The order of magnitude capital costs and net annual costs are calculated for the following size mills which are considered typical for the industry:

Kraft	500 ADT/Day
Sulfite	200 ADT/Day
NSSC	200 ADT/Day

Following the same procedure as Chapter 5, costs for "site clearance" and "loss of production" are not included in any calculations in this chapter.

6.2 GENERAL DESCRIPTION OF CONTROL METHODS

For a general description of control equipment, the reader should refer to Chapter 5. The general descriptions included in this chapter refer to "New Developments."

While capital costs for the new recovery designs were included in Chapter 5, a detailed discussion of the two new systems (High Solids and ACE) are included in this chapter. These two new systems are classified as new developments; however, the initial installations of these systems have just recently been placed in operation (start-up was the last six months of 1969).

Operating data from these first two systems were not available, and, therefore, could not be included with this report.

6.2.1 CONVENTIONAL KRAFT RECOVERY SYSTEMS

The following new developments are applicable to a conventional kraft recovery system with a direct contact evaporator.

6.2.1.1 Black Liquor Oxidation

Supplementing the discussion of oxidation systems in Section 5.3.6.1.5, the West Coast Canadian Kraft Industry reportedly may sponsor a pilot plant study of a recovery flue gas scrubbing system using a sodium carbonate-bicarbonate mixture at pH 9.4 and 180°F. However, the system design must also include an elaborate oxidation system to oxidize the Na_2S formed since the liquor is recirculated to reduce the chemical cost. Until more details of this work done by British Columbia Research Council are available, no evaluations of the scrubbing liquor and the system can be made.

6.2.1.2 "Hot" Precipitators

The question has been raised as to whether the so-called "Hot" precipitators would be of any interest to the U.S. pulp industry. These precipitators have found acceptance in Scandinavia.

The most common location is downstream of the economizer--"warm" precipitator--but in at least twenty-five mills, the precipitator is located before the economizer--"hot" precipitator.

The hot precipitator provides many advantages for the economizer. Less cleaning is necessary. The cleaning is done by flushing the economizer tubes with water. For the warm precipitator, the maintenance cost of the economizer is high even when using hot cleaning water. The reduced cleaning also means a longer life of the economizer. The hot precipitator will reduce the solids in the flue gas, which will keep the heat transfer coefficient at the high level as well as the draft loss constant through the economizer. According to the Scandinavian experience, the economizer can be made approximately 30 percent smaller when the precipitator is located upstream (hot) due to constant heat transfer coefficient.

The resistivity of the dust at the elevated temperature is considerably lower than for conventional precipitator operation. This would mean that the size of the precipitator could be smaller.

The gas volume at this temperature is, however, much larger and the end result is that the hot precipitator has to be made larger than the warm unit. The savings in the smaller economizer will be offset by the larger precipitator.

The layout is somewhat simpler for the hot concept, which will realize some savings, especially in Scandinavia, where most precipitator installations are inside the building. The structural steel, however, has to be increased due to the increased load. The hot precipitator is larger and heavier and is always located above the boiler. If both precipitator designs are properly maintained, the total evaluated cost of the "warm" precipitator is approximately equal to the total evaluated cost of the "hot" precipitator.

The high temperature creates more severe operating conditions, and the dust has more sticky properties than for the warm precipitator. For these reasons, this control method appears to have limited application in the U.S.A.

6.2.1.3 Precipitator Installed Subsequent to Venturi Evaporator/Scrubber

There has been some discussion about adding an electrostatic precipitator following a Venturi evaporator/scrubber on a recovery boiler. This is a somewhat unorthodox approach, and there is knowlingly none in operation, but one is to start up in 1970 in the aluminum industry. According to the manufacturer, the corrosion problem anticipated is overcome by making the precipitator shell in tile, the collecting plates in treated wood, and all other internals in 316 stainless steel. Further, there are no rapping mechanisms because the manufacturer expects that considerable condensation will take place on the collector plates which will flush down the precipitated dust. The precipitator is further equipped with wet bottom. The equipment cost is estimated to be approximately 30 percent higher than for a conventional precipitator installed on the same recovery boiler.

Operating experience from this first unit will tell if this approach is feasible. At the present time, this control method must be classified as questionable.

6.2.2 NEW KRAFT RECOVERY SYSTEM

This section of the report describes the conversion of existing recovery systems to the new designs which remove the direct contact evaporator from the flue gas stream. These designs for new installations were generally described in Section 5.3.6.1.6 with capital costs.

These new designs have a number of variations; however, they may be generally classified into the following two groups:

1. High Solids Concentration in the Multiple Effect Evaporators. (High Solids System)

2. Air Contact Evaporator System (ACE)

More detailed analyses of the various reduced odor systems have been reported by Arhippainen (3), Hochmuth (1), and Roberson (4).

The description of these recovery systems is discussed as follows:

High Solids System
ACE System
Multiple Effect Evaporators for High Solids or ACE Systems
Precipitators

6.2.2.1 High Solids Systems

This system is so called because the black liquor solids leaving the multiple effect evaporators are in the range of 60 percent to 70 percent solids. Currently, the most acceptable solids concentration is expected to be about 63 percent. The liquor is then heated by direct steam heaters and fired into the recovery furnace. (Figure 6-1)

By concentrating the liquor to a firing solids concentration of 63 percent, the conventional direct contact evaporator may be removed from the flue gas stream. The economizer surface is increased to reduce the flue gas temperature from approximately 600°F leaving the boiler to 350° - 400°F leaving the economizer. For an existing unit, the flue gas temperature must be below 400°F. For an existing unit, the flue gas temperature must be below 400°F if a tile-shell precipitator is on line. This can usually be accomplished for most installations. However, in cases of high temperature feed-water (in the order of 350°F and higher), this gas temperature may be difficult to attain.

The high solids system has found extensive application in Europe due to the high fuel cost. These systems were originally installed in Europe to attain high thermal economy.

6.2.2.2 Air Contact Evaporator System

This system has been described in more detail by Hockmuth (1). It consists essentially of removing the conventional cascade evaporator from the flue gas stream and placing it in the heated air stream as shown in Figure 6-2. In order to reduce the temperature of the flue gases to a reasonable level a regenerative type air heater is added.

The flue gases from the boiler at approximately 650°F, pass to the air heater and are reduced to approximately 300 - 350°F. The heat removed from the flue gases is transferred to the combustion air which is heated from 80°F to approximately 550°F. The heated combustion air then passes through the air contact evaporator where the temperature is reduced to approximately 275°F before being introduced into the furnace.

As the air temperature drops from 550°F to 275°F, the black liquor is concentrated from 55 percent solids to approximately 65 percent solids by evaporation. Any odorous gases which are entrained with the combustion air are oxidized to non-odorous compounds in the furnace.

For the best thermal economy, the black liquor is usually concentrated to 55 percent solids in the multiple effect evaporators.

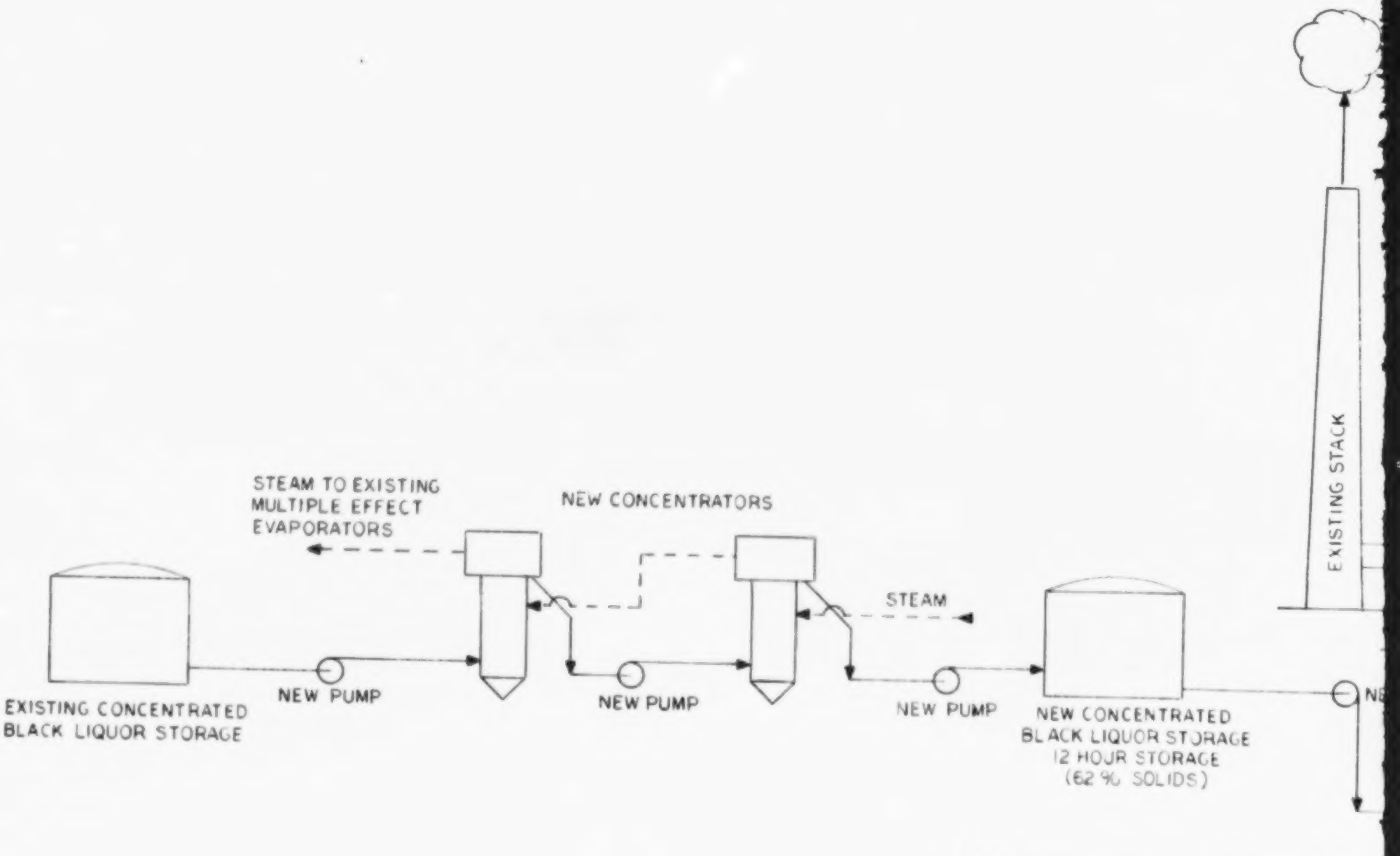
The application of the ACE system to existing units will usually require increasing the building size. Therefore, the cost for site clearance will probably be very expensive. In many cases, it may well be impossible to install the ACE system at an existing mill.

6.2.2.3 Multiple-Effect Evaporators

The multiple-effect evaporators at existing mills are usually designed to produce 50 percent black liquor solids.

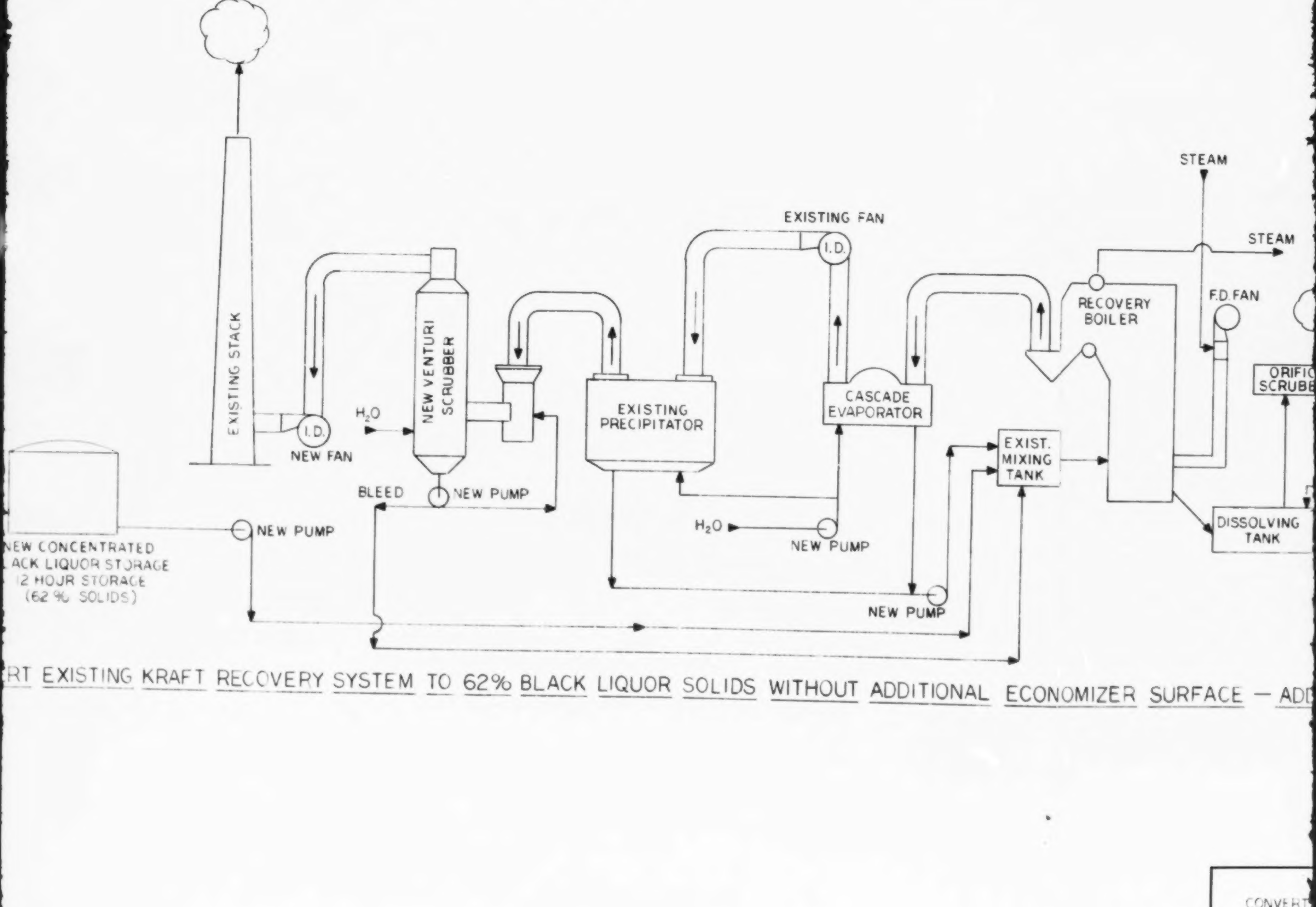
If the High Solids System is installed at an existing mill, an additional effect, usually referred to as a "concentrator," is added to provide 63 percent solids liquor. The flash steam from the concentrator can usually be introduced into the 3rd or 4th evaporator effect. As shown in Figure 6-2, the additional effect is a two-body installation.

If the ACE System is installed at an existing mill, an additional effect must be added to concentrate the liquor to 55 percent. For a new installation, this can be incorporated into the original design of the multiple-effect evaporators. The flash steam from


STEAM TO EXISTING MULTIPLE EFFECT EVAPORATORS
NEW CONCENTRATORS
EXISTING STACK
STEAM
EXISTING CONCENTRATED BLACK LIQUOR STORAGE
NEW PUMP
NEW PUMP
NEW PUMP
NEW CONCENTRATED BLACK LIQUOR STORAGE 12 HOUR STORAGE (62 % SOLIDS)
NE

CONVERT EXISTING KRAFT RECO

A


EXISTING STACK
I.D.
NEW FAN
H₂O
NEW VENTURI SCRUBBER
BLEED
NEW PUMP
EXISTING PRECIPITATOR
EXISTING FAN
I.D.
CASCADE EVAPORATOR
H₂O
NEW PUMP
NEW PUMP
EXIST. MIXING TANK
STEAM
STEAM
RECOVERY BOILER
F.D. FAN
ORIFIC SCRUBB
DISSOLVING TANK
NEW PUMP
NEW CONCENTRATED
ACK LIQUOR STORAGE
12 HOUR STORAGE
(62 % SOLIDS)

RT EXISTING KRAFT RECOVERY SYSTEM TO 62% BLACK LIQUOR SOLIDS WITHOUT ADDITIONAL ECONOMIZER SURFACE — ADD

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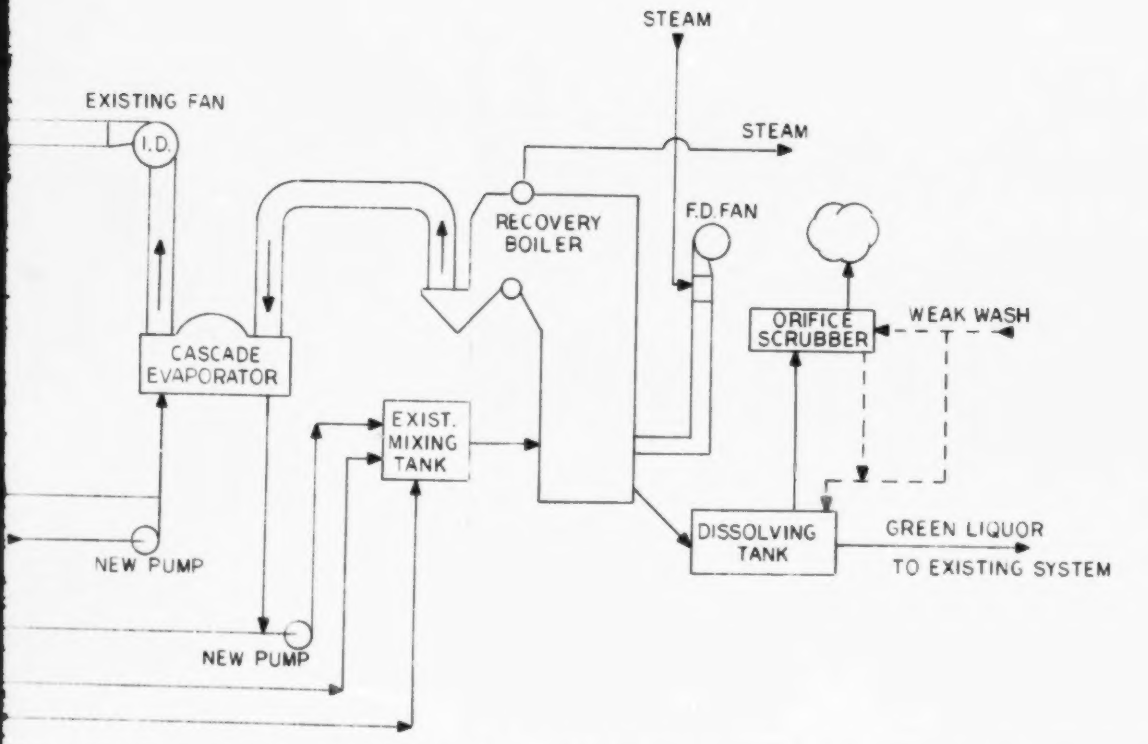
STEAM
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STEAM
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BOILER
F.D. FAN
ORIFICE
SCRUBBER
WEAK WASH
CASCADE
EVAPORATOR
EXIST.
MIXING
TANK
NEW PUMP
DISSOLVING
TANK
GREEN LIQUOR
TO EXISTING SYSTEM
NEW PUMP

DS WITHOUT ADDITIONAL ECONOMIZER SURFACE — ADD VENTURI SCRUBBER

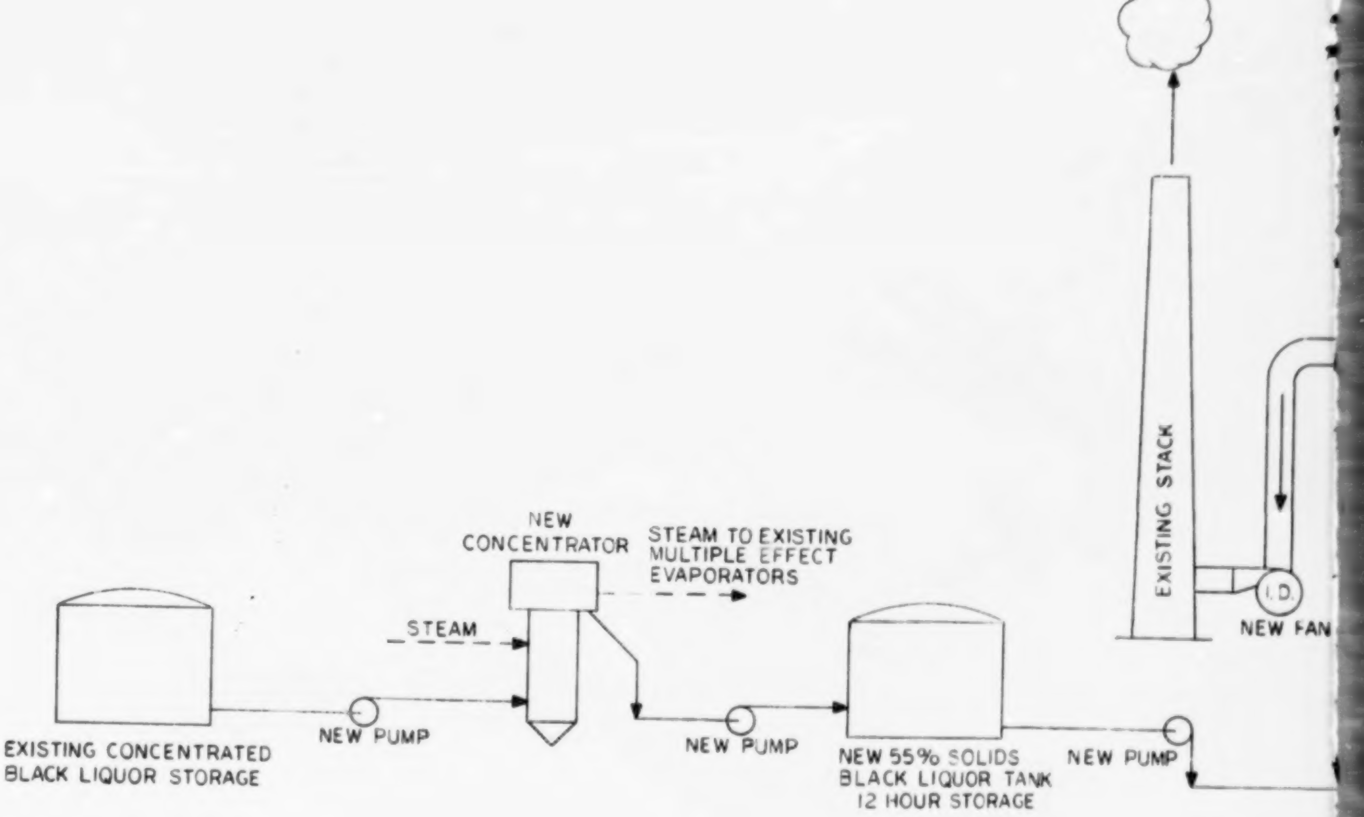
H.J.S.

SCHEMATIC ARRANGEMENTS CONVERT EXISTING RECOVERY TO 62% BLACK LIQUOR SOLIDS USE EXISTING ECONOMIZER AND ADD VENTURI SCRUBBER	EXHIBIT NO. FIG. 6-1

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

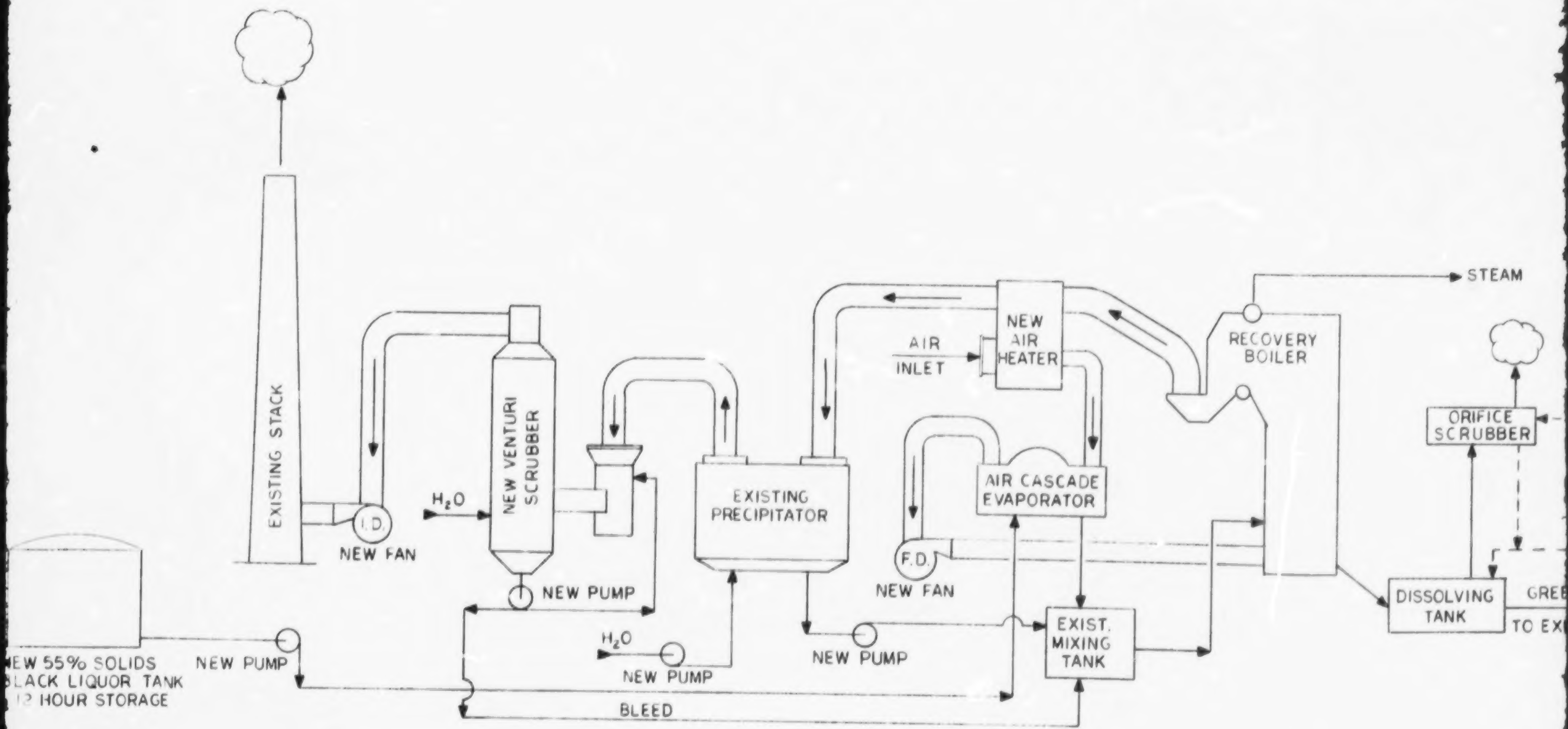
ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


EXISTING STACK
I.D.
NEW FAN
NEW CONCENTRATOR
STEAM TO EXISTING MULTIPLE EFFECT EVAPORATORS
STEAM
EXISTING CONCENTRATED BLACK LIQUOR STORAGE
NEW PUMP
NEW PUMP
NEW 55% SOLIDS BLACK LIQUOR TANK 12 HOUR STORAGE
NEW PUMP

CONVERT EXISTING KRAFT REC

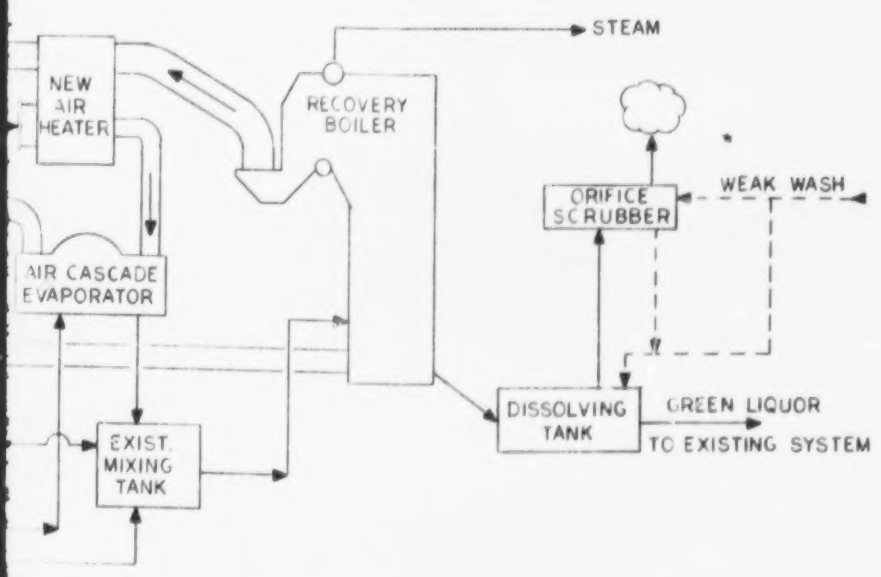
A


EXISTING STACK
NEW VENTURI SCRUBBER
I.D.
NEW FAN
H_2O
NEW PUMP
H_2O
NEW PUMP
EXISTING PRECIPITATOR
NEW PUMP
STEAM
NEW AIR HEATER
AIR INLET
RECOVERY BOILER
AIR CASCADE EVAPORATOR
F.D.
NEW FAN
EXIST. MIXING TANK
ORIFICE SCRUBBER
DISSOLVING TANK
GREE
TO EX
EW 55% SOLIDS
LACK LIQUOR TANK
12 HOUR STORAGE
NEW PUMP
BLEED

CONVERT EXISTING KRAFT RECOVERY SYSTEM TO AIR CASCADE EVAPORATOR AND ADD VENTURI SCRUBBER

B

C


STEAM
NEW AIR HEATER
RECOVERY BOILER
ORIFICE SCRUBBER
WEAK WASH
AIR CASCADE EVAPORATOR
DISSOLVING TANK
GREEN LIQUOR
TO EXISTING SYSTEM
EXIST. MIXING TANK

OR AND ADD VENTURI SCRUBBER

HJS

SCHEMATIC ARRANGEMENTS
CONVERT EXISTING RECOVERY TO AIR CASCADE EVAPORATOR
AND ADD VENTURI SCRUBBER

EXHIBIT NO.
FIG. 6-2

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

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J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

the additional effect can usually be exhausted into the existing evaporators.

The exact design for increasing the capacity of the multiple-effect evaporators at an existing mill to provide 55 percent or 63 percent solids will depend upon the particular situation, and will have to be individually investigated.

In general, 63 percent solids is currently used for evaporator design. Higher solids concentration (above 63 percent) are expected to require higher evaporator capital costs and operating costs. However, more experience is required before the optimum solids concentration can be determined.

The first U.S.A. installation for 63 percent solids has been reported by Groce and Harris (2). This installation is for a new recovery system at an existing mill.

6.2.2.4 Precipitators

Precipitators have been discussed in detail in Chapter 5. The cost data for Chapter 5 are based on a precipitator for a conventional recovery system which usually has a particulate loading in the order of 2 grains per ACF. For these new recovery systems which omit the direct contact evaporator, the particulate loading will approximately double to 4 grains per ACF. This is attributed to the fact that the direct contact evaporator usually removes approximately 50 percent of the entering particulate.

This higher particulate loading of 4 grains per ACF means that the precipitator for a given size recovery system will have to be larger and slightly more expensive for the new recovery systems. Therefore, a precipitator for a new recovery system will have a capital cost approximately 10 percent higher than a precipitator for a conventional recovery system in order to achieve the same particulate loading leaving either precipitator.

As discussed in Section 6.2.1.2, Hot Precipitators have also been applied to high solids recovery systems in Scandinavia.

6.2.3 SULFITE SOURCES

Sulfite control methods which are considered in this chapter include packed towers which have been described in detail in Chapter 5.

6.2.4 NSSC Sources

Due to the lack of emission data (as described in Chapter 5), control methods for NSSC sources were not analyzed.

6.3 APPLICATION, COST, AND EFFECTIVENESS OF NEW DEVELOPMENTS IN CONTROL TECHNOLOGY

As stated in the introduction of this chapter, the capital costs and net annual costs for this chapter should be considered order of magnitude values which will provide an approximate basis of comparison for the different control methods.

Following the same procedure as Chapter 5, costs for "site clearance" and "loss of production" are not included in any calculations in this chapter.

6.3.1 KRAFT SOURCES

The following new developments were evaluated for kraft sources:

Source	Description
Recovery System	Case 1 - The conversion of an existing recovery system with a conventional direct contact evaporator (precipitator located on the ground) to the following new systems: a. High solids with large economizer. b. ACE System. c. High solids with no change to the existing economizer.
Recovery System	Case 2 - Same as Case I except the existing precipitator is located on the building roof.
Recovery System	Case 3 - The conversion of an existing Venturi scrubber to the following new system: a. High solids system with precipitator. b. Wet precipitator and black liquor oxidation. c. Brine system.

Source	Description
Recovery System	Case 4 - Molecular oxidation system.
Oxidation System	Analysis of new developments to control emissions from weak and concentrated oxidation systems.
Smelt Dissolving Tanks	Analysis of new developments to control emissions from smelt dissolving tanks.
Digester Relief and Blow, Multiple Effect	Analysis of an individual thermal oxidation system.
Combination Boilers	Analysis of mechanical collectors with shave-off scrubbers.
Brown Stock Washer Vents	Analysis of continuous diffusion washing, enclosed pressure washing, and incineration of vent gases.
Multiple Effect Evaporators	Discussion of caustic scrubbers.
Miscellaneous	Discussion of steam stripping of combined condensate.

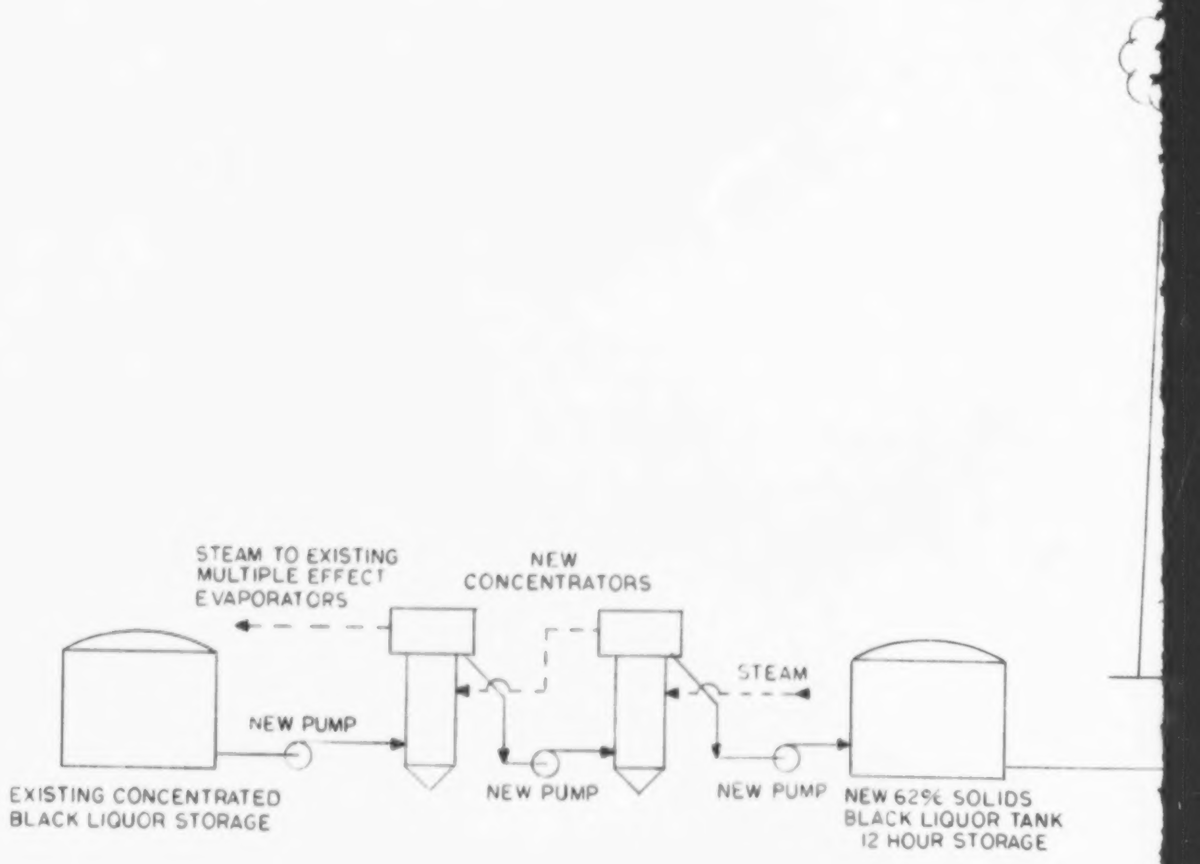
6.3.1.1 Recovery System, Case 1

Application

This case involves the conversion of an existing recovery boiler with a 90 percent AOE precipitator, which is located on the ground, to a new design which eliminates the direct contact of the flue gas and black liquor. This case also includes the addition of a new 99.0 percent guaranteed efficiency Venturi scrubber following the existing precipitator. Three different types of control methods were investigated as follows:

a. Conversion of the existing recovery system to an ACE System, and adding a concentrator to obtain 55 percent solids liquor. (Figures 6-2, 6-5, and 6-6)

b. Conversion of the existing recovery system to a high solids system (62 percent solids) and adding additional economizer surface. (Figures 6-3 and 6-4)


STEAM TO EXISTING MULTIPLE EFFECT EVAPORATORS
NEW CONCENTRATORS
STEAM
NEW PUMP
NEW PUMP
NEW PUMP
EXISTING CONCENTRATED BLACK LIQUOR STORAGE
NEW 62% SOLIDS BLACK LIQUOR TANK 12 HOUR STORAGE

CONVERT EXISTING KRAFT RECOVERY SYSTEM TO 62% BLACK LIQUO

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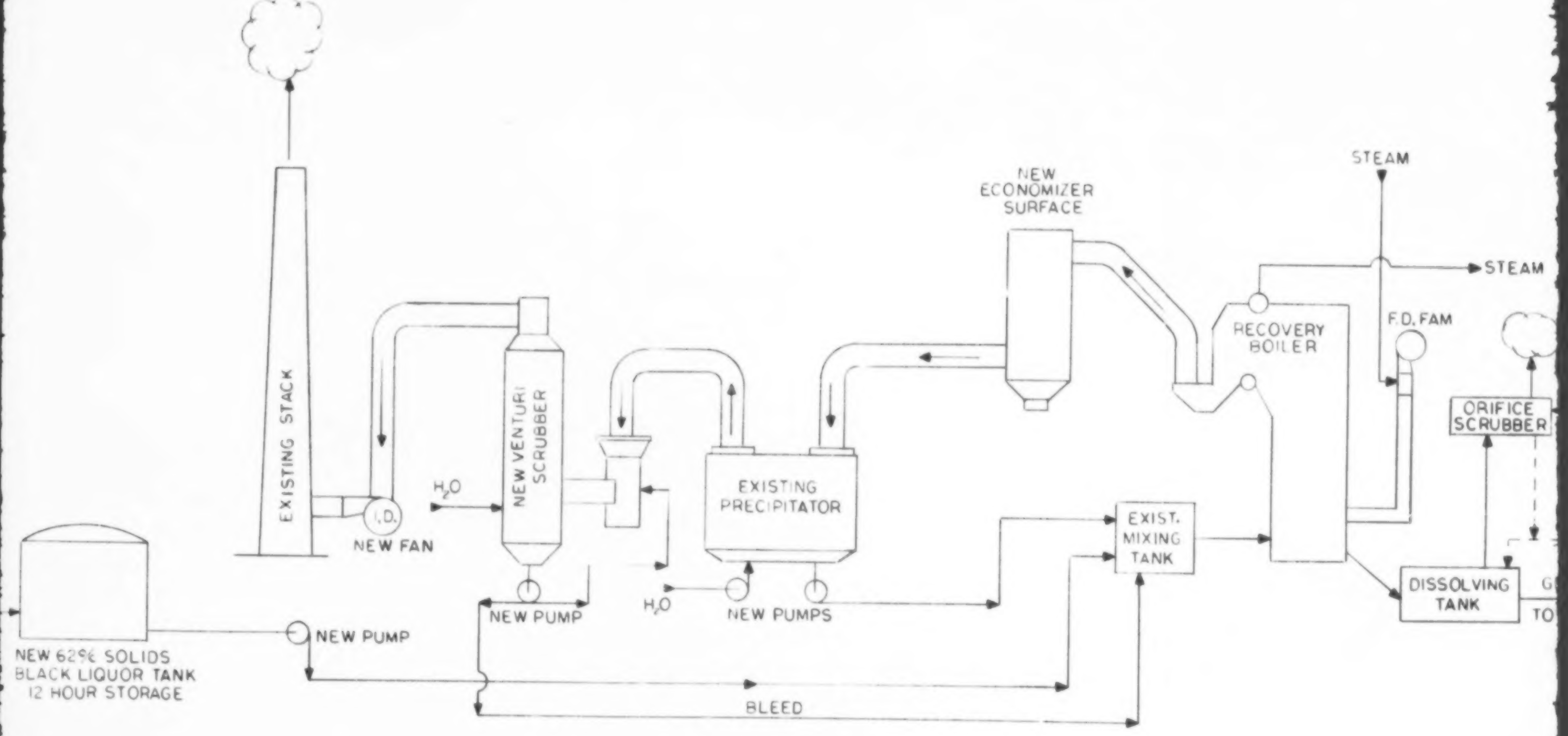
EXISTING STACK
I.D.
NEW FAN
H2O
NEW VENTURI SCRUBBER
NEW PUMP
EXISTING PRECIPITATOR
H2O
NEW PUMPS
NEW ECONOMIZER SURFACE
STEAM
STEAM
RECOVERY BOILER
F.D. FAM
ORIFICE SCRUBBER
EXIST. MIXING TANK
DISSOLVING TANK
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TO
NEW 62% SOLIDS BLACK LIQUOR TANK 12 HOUR STORAGE
NEW PUMP
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0 62% BLACK LIQUOR SOLIDS WITH ADDITIONAL SURFACE AND ADD VENTURI SCRUBBER

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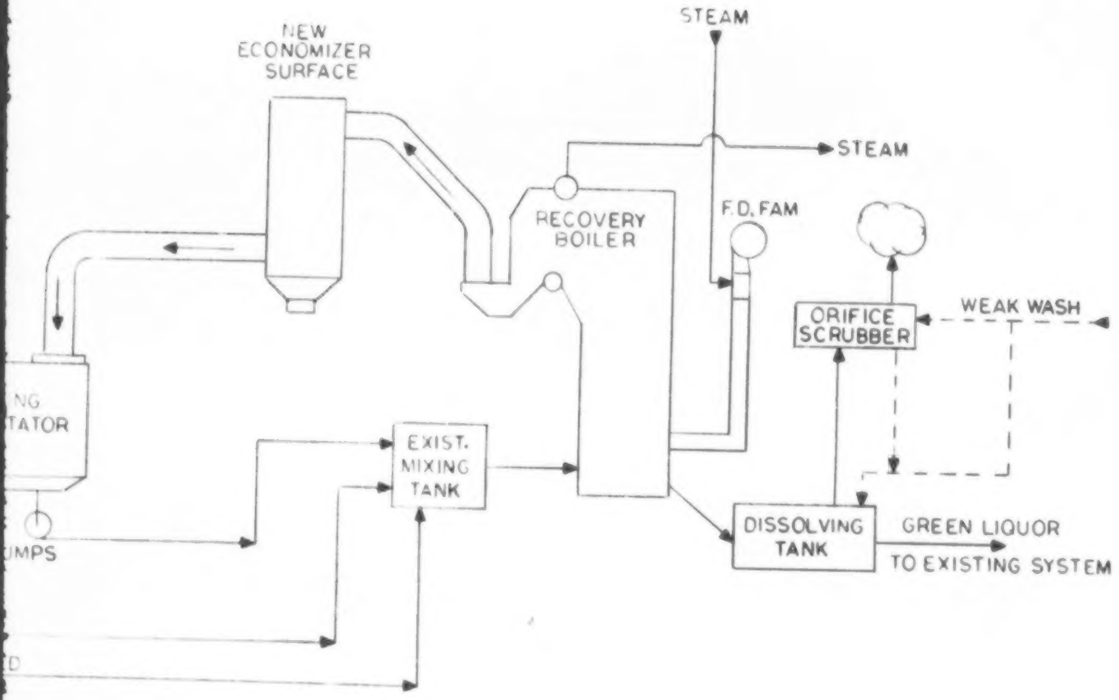
NEW ECONOMIZER SURFACE
STEAM
STEAM
RECOVERY BOILER
F.D. FAM
ORIFICE SCRUBBER
WEAK WASH
NG
TATOR
EXIST. MIXING TANK
UMPS
D
DISSOLVING TANK
GREEN LIQUOR
TO EXISTING SYSTEM

D ADD VENTURI SCRUBBER

HJS

SCHEMATIC ARRANGEMENT
CONVERT EXISTING RECOVERY SYSTEM TO 62% BLACK LIQUOR SOLIDS AND ADD ADDITIONAL ECONOMIZER SURFACE

EXHIBIT NO.
FIG. 6-3

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
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FOR
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CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
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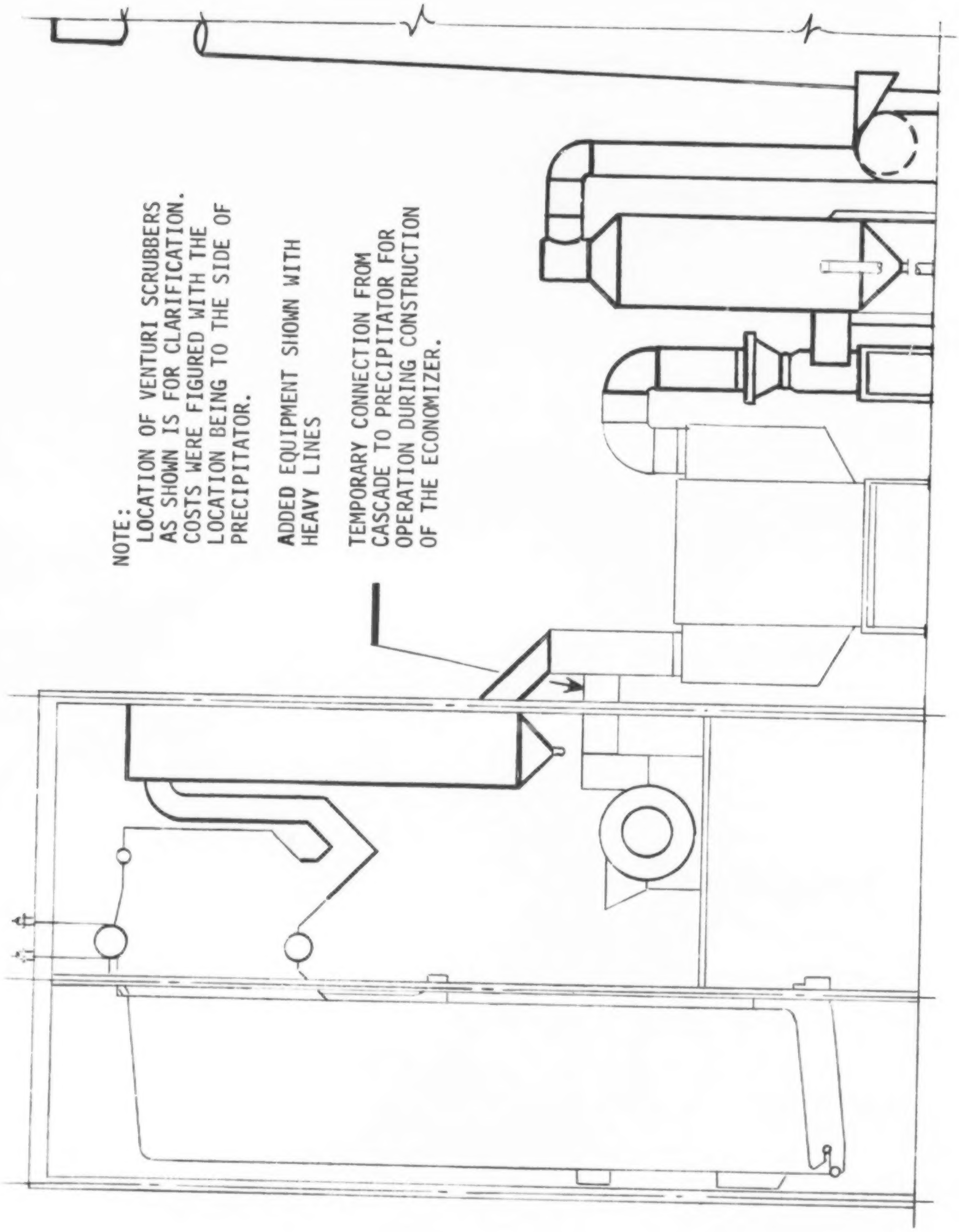
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.

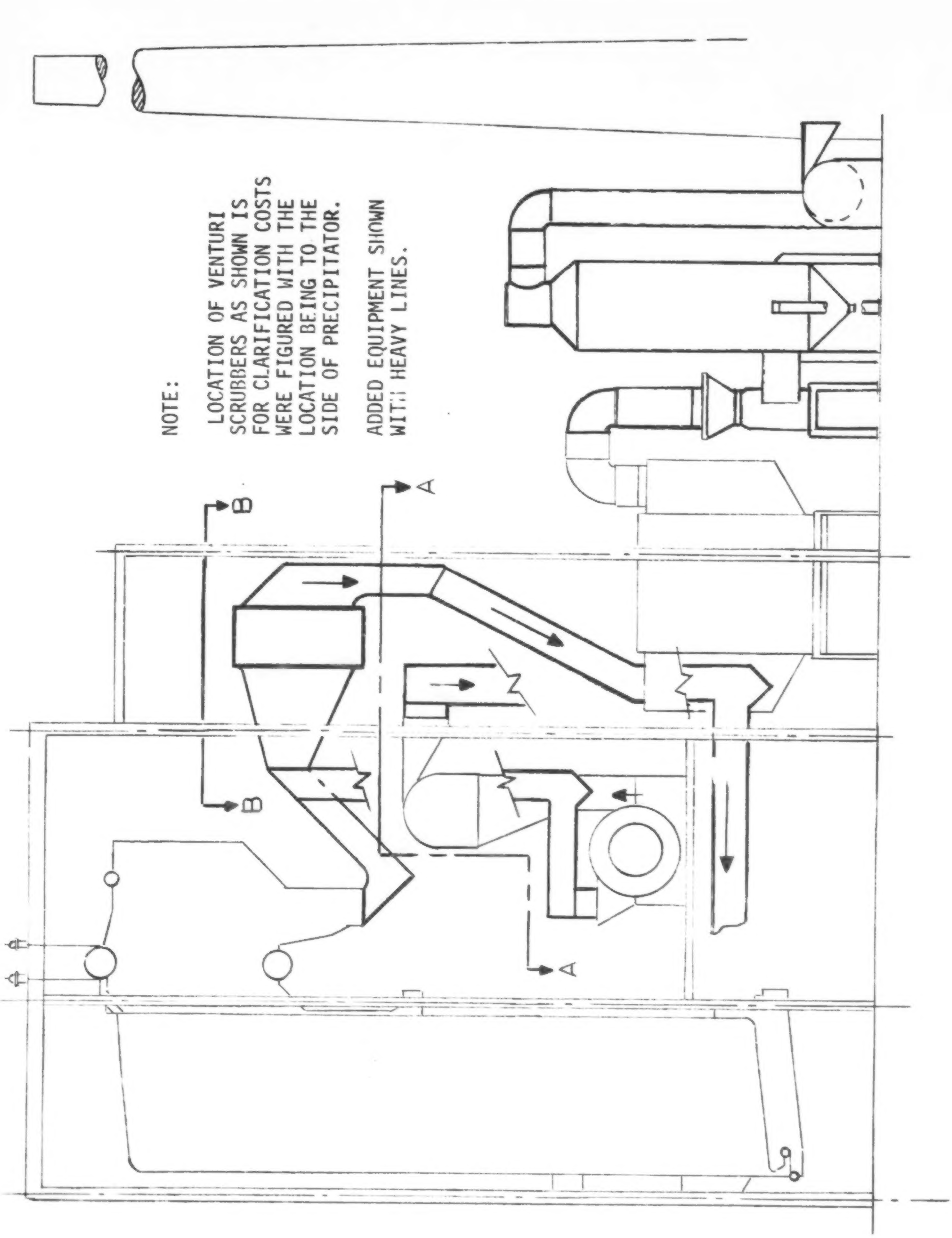


NOTE:
LOCATION OF VENTURI SCRUBBERS AS SHOWN IS FOR CLARIFICATION. COSTS WERE FIGURED WITH THE LOCATION BEING TO THE SIDE OF PRECIPITATOR.

ADDED EQUIPMENT SHOWN WITH HEAVY LINES

TEMPORARY CONNECTION FROM CASCADE TO PRECIPITATOR FOR OPERATION DURING CONSTRUCTION OF THE ECONOMIZER.

FIG. 6-4 TYPICAL ARRANGEMENT FOR CONVERSION TO 62% BLACK LIQUOR RECOVERY SYSTEM, LARGE ECONOMIZER, AND VENTURI SCRUBBER ADDITION - RECOVERY BOILER



NOTE:

LOCATION OF VENTURI SCRUBBERS AS SHOWN IS FOR CLARIFICATION COSTS WERE FIGURED WITH THE LOCATION BEING TO THE SIDE OF PRECIPITATOR.

ADDED EQUIPMENT SHOWN WITH HEAVY LINES.

FIG. 6-5 TYPICAL ARRANGEMENT FOR CONVERSION TO AIR CASCADE EVAPORATOR AND VENTURI SCRUBBER ADDITION TO AN EXISTING PRECIPITATOR - RECOVERY BOILER

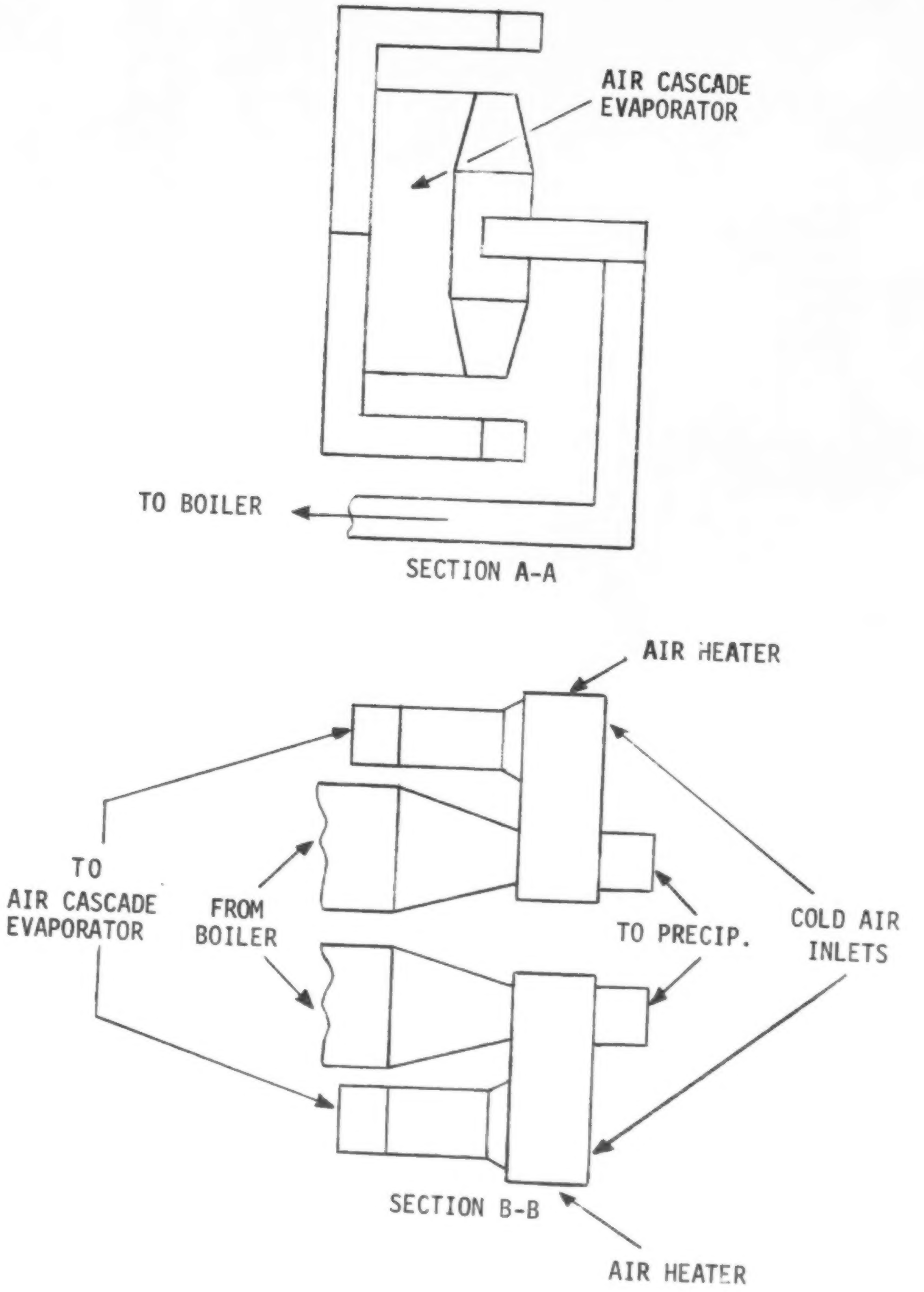

AIR CASCADE EVAPORATOR
TO BOILER
SECTION A-A
AIR HEATER
TO AIR CASCADE EVAPORATOR
FROM BOILER
TO PRECIP.
COLD AIR INLETS
SECTION B-B
AIR HEATER

FIG. 6-6 TYPICAL ARRANGEMENT FOR CONVERSION TO AIR CASCADE EVAPORATOR AND VENTURI SCRUBBER ADDITION TO AN EXISTING PRECIPITATOR - RECOVERY BOILER

c. Conversion of the existing recovery system to a high solids system (62 percent solids) with no change to the economizer surface. (Figure 6-1)

This last method sacrifices thermal efficiency for a lower capital cost and a reduced down time for converting the recovery unit. As shown in Figure 6-1, water is added to the cascade evaporator in the last control method (c) in order to reduce flue gas temperature to a more reasonable level (below 400°F for tile shell precipitators) before entering the existing precipitator. This temperature limit does not apply to steel shell precipitators; however, there are very few steel units in operation. The flue gas temperature may also have to be controlled to remain within the design gas flow volumes for the precipitator and the I.D. fan.

The reduction of the flue gas temperature by the addition of water to the cascade is considered more reasonable than spraying water directly into the flue gas stream. If water is sprayed directly into the flue gas stream, some water droplets may reach the precipitator resulting in significant corrosion problems.

This reduction in flue gas temperature by the addition of moisture will increase the visibility of the plume being emitted from the stack for control method (c). Conversely, methods (a) and (b) should give a less visible plume. As mentioned previously, all three of these control methods include the addition of a 99.0 percent guaranteed efficiency Venturi scrubber added downstream of the existing precipitator.

The selection of the most effective control system will depend upon a number of factors which must be evaluated individually for each specific application. For example, additional space will be required for all three control methods. Further, the electrical requirements will increase, especially for the ACE system. This may be an important consideration for some mills.

Costs (500 Ton Size)

Capital cost of these three control systems are as follows:

Control Method	Capital Cost
a. Air Contact Evaporator	$ 1,500,000
b. 62% B.L.Solids with Additional Economizer Surface	1,000,000
c. 62% B.L.Solids, without Additional Economizer Surface	500,000

The net annual costs for these three control methods are as follows:

Control Method	Net Annual Cost
a. Air Contact Evaporator	400,000
b. 62% B.L.Solids with Additional Economizer Surface	275,000
c. 62% B.L.Solids without Additional Economizer Surface	200,000

The net annual cost includes credit for chemical recovery, resulting from the collection of additional salt cake by the Venturi scrubber. The salt cake collected by the Venturi scrubber will be bled into the existing salt cake mix tank.

Effectiveness

Particulate. The particulate removal for all three cases is considered to be approximately equally effective. Actually the control method utilizing 62 percent solids liquor with no change to the economizer would be slightly more effective due to the fact that the cascade evaporator would have been retained and would collect approximately 50 percent of the inlet dust.

Sulfur Dioxide and Total Sulfur Removal. The emission of SO_2 and TRS is considered to be equally low for all three control methods.

Operation. Maintenance and operating costs are not well defined at this time since none of these control methods has been in operation for sufficient time to clearly establish these costs.

However, if we can assume that higher maintenance and operating costs are usually incurred with more equipment, as evidenced by significant differences in capital cost, then the maintenance and operating costs will be in the following relationship:

Highest - ACE System

Moderate - High Solids System, large economizer

Lowest - High Solids System, no change to the economizer

The reliability of all three control methods has not been clearly defined at this time, but is expected to be reasonably equal. The high solids system utilizing 62 percent solids with no change to the economizer does have one disadvantage as compared to the other two methods. This is in regard to plume appearance. Since additional moisture must be used to reduce the flue gases leaving the economizer from approximately 600° F to 300-350° F, the moisture content of the flue gases will be higher than the other two methods. This will result in a more visible plume leaving the stack as compared to methods (a) and (b). The plume visibility for method (c) would be approximately equal to a conventional recovery with a direct contact evaporator.

Summary

If only these three control methods are compared, control method (c) has the lowest net annual cost with an effectiveness equal to the other two control methods, as described previously.

Comparison with Chapter 5. If this control method is compared to concentrated black liquor oxidation and the addition of a Venturi scrubber (from Chapter 5), the following net annual costs are applicable:

	Net Annual Cost
Concentrated B.L. Oxidation (99% Oxidation Efficiency)	$ 75,000 (from Fig. 5-40)
99.0% Guaranteed Efficiency Venturi Scrubber	100,000 (from Fig. 5-21)
Total	$ 175,000

This net annual cost of $175,000 compares to $200,000 for the 62 percent solids without additional economizer surface. Considering the fact that oxidation will require less downtime for conversion, the concentrated oxidation system appears to be the best choice for an existing installation. The oxidation system will probably have a somewhat higher malodorous emission than the 62 percent solids. For example, the relative difference in H_2S concentration ranges in the stack gases might be approximately 5-20 ppm for oxidation and 1-10 ppm for 62 percent solids, when these control methods are applied to recovery systems of moderate age (within the last 10 years).

6.3.1.2 Recovery System, Case 2

Application

Case 2 considers the same three control methods as Case 1 with the exception that the precipitator is located on the roof, and has an AOE of 95 percent.

Costs

All the considerations for Case 1 apply equally to Case 2. The only difference is that the capital cost and net annual costs will change. Whether the change is an increase or a decrease is not significant since the order of magnitude will be relatively small.

Effectiveness

The effectiveness comparison for this case will be the same as Case 1.

6.3.1.3 Recovery System, Case 3

Application

Case 3 is based on reducing emissions from an existing single stage Venturi evaporator/scrubber. The existing single stage Venturi is assumed to have a particulate removal efficiency of 80 percent AOE. The following control methods have been investigated:

a. Convert to High Solids Systems, omitting contact between the flue gas and black liquor, and add precipitator.

b. Add new design wet precipitator (see paragraph 6.2.1.3) following Venturi and install new oxidation system.

c. Install Na_2SO_4 Brine System.

These control methods are described in more detail as follows:

a. Convert to High Solids System, Omitting Contact Between the Flue Gas and Black Liquor and Add Precipitator. For this control method, the existing Venturi will be removed and the cyclone separator converted to a cyclone evaporator which utilizes water rather than black liquor. The cyclone evaporator is retained for several reasons: 1) to reduce the gas volume to within fan design limits, 2) to reduce the gas temperature to within fan and precipitator design limits, and 3) to provide some particulate collection which will reduce slightly the cost of the new precipitator.

Therefore, the flue gas temperature will now be in the order of 350°F rather than 180°F to prevent corrosion problems in the precipitator. The salt cake collected in the cyclone evaporator will be concentrated to a 65 percent solids slurry of water and salt cake. The necessary bleed-off to maintain this solids level will flow to the existing salt cake mixing tank.

A new precipitator will be added at ground level to achieve minimum particulate emission. The required removal efficiency of the precipitator may vary depending upon local regulations. However, for the purpose of this control method the maximum of 99.9 percent guarantee has been included. This results in an AOE of 99.5 percent.

This control method provides minimum particulate emission (99.9 percent guarantee precipitator) and minimum TRS (total reduced sulfur) emission. The minimum TRS emission is achieved by eliminating the contact of the flue gas with the black liquor.

b. Add New Design Wet Precipitator Following Venturi and Install New Oxidation System. One precipitator designed for 200°F and following a single stage Venturi is currently being installed in the aluminum industry (see paragraph 6.2.1.3). Therefore, this type design has been investigated as a possible control method. The materials of construction are as follows:

Shell - Tile

Collecting Plates - Treated Wood

Internals - 316L Stainless Steel

There is no rapping system since considerable condensation will take place on the collector plates. The manufacturer expects that this condensation will flush down the precipitated dust. The ductwork entering and leaving the precipitator is 316 stainless. A new 250 foot brick chimney with an acid brick lining is included.

The precipitator for this method has a guaranteed efficiency of 99.5 percent (99.0 percent AOE) for a combined efficiency of 99.8 percent AOE which includes the 80 percent Venturi AOE.

A concentrated black liquor oxidation system at 99 plus percent efficiency is included to minimize TRS emission.

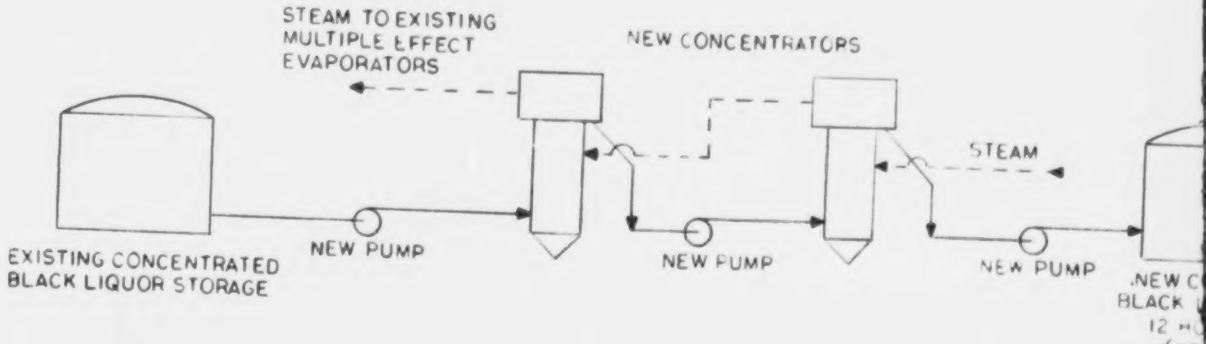
c. Install Na_2SO_4 Brine System. Another control method which was considered was converting a Venturi evaporator/scrubber to a 15 percent Na_2SO_4 brine Venturi unit as shown in figure 6-7. This method also eliminates the direct contact of the flue gas and the black liquor to achieve minimum TRS emission. The solids concentration in the brine solution is limited to 15 percent to achieve an expected particulate collection efficiency of 99 percent. The Venturi pressure drop will be increased to 30 inches to achieve this 99 percent efficiency.

As shown in figure 6-7, the bleed from the brine system is concentrated to 65 percent solids and then introduced into the liquor system at the salt cake mix tank.

Costs (500 Ton Size)

Based on the above descriptions, order of magnitude capital costs and net annual costs are as follows:

Control Method	Capital Cost	Net Annual Cost
a. High Solids	$ 1,800,000	$ 320,000
b. Wet Precipitator	$ 1,000,000	$ 150,000
c. Na_2SO_4 Brine System	$ 1,500,000	$ 330,000


STEAM TO EXISTING
MULTIPLE EFFECT
EVAPORATORS
NEW CONCENTRATORS
STEAM
EXISTING CONCENTRATED
BLACK LIQUOR STORAGE
NEW PUMP
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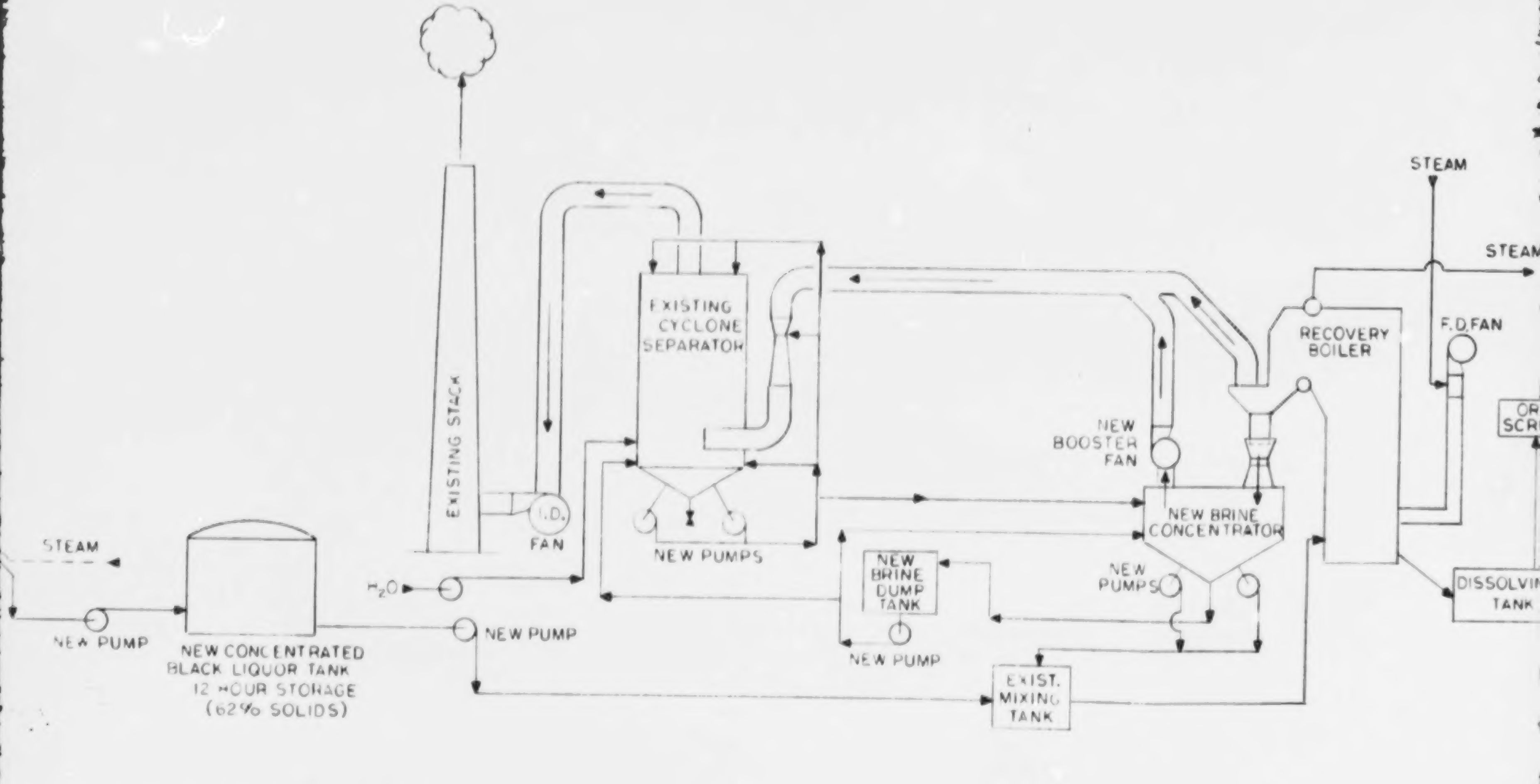
STEAM
NEW PUMP
NEW CONCENTRATED BLACK LIQUOR TANK 12 HOUR STORAGE (62% SOLIDS)
EXISTING STACK
I.D. FAN
H_2O
NEW PUMP
EXISTING CYCLONE SEPARATOR
NEW PUMPS
NEW BRINE DUMP TANK
NEW PUMP
EXIST. MIXING TANK
NEW BOOSTER FAN
NEW BRINE CONCENTRATOR
NEW PUMPS
RECOVERY BOILER
STEAM
STEAM
F.D.FAN
OR
SCR
DISSOLVIN
TANK

LIQUOR VENTURI RECOVERY SYSTEM TO 15% BRINE VENTURI AND 62% BLACK LIQUOR SOLIDS

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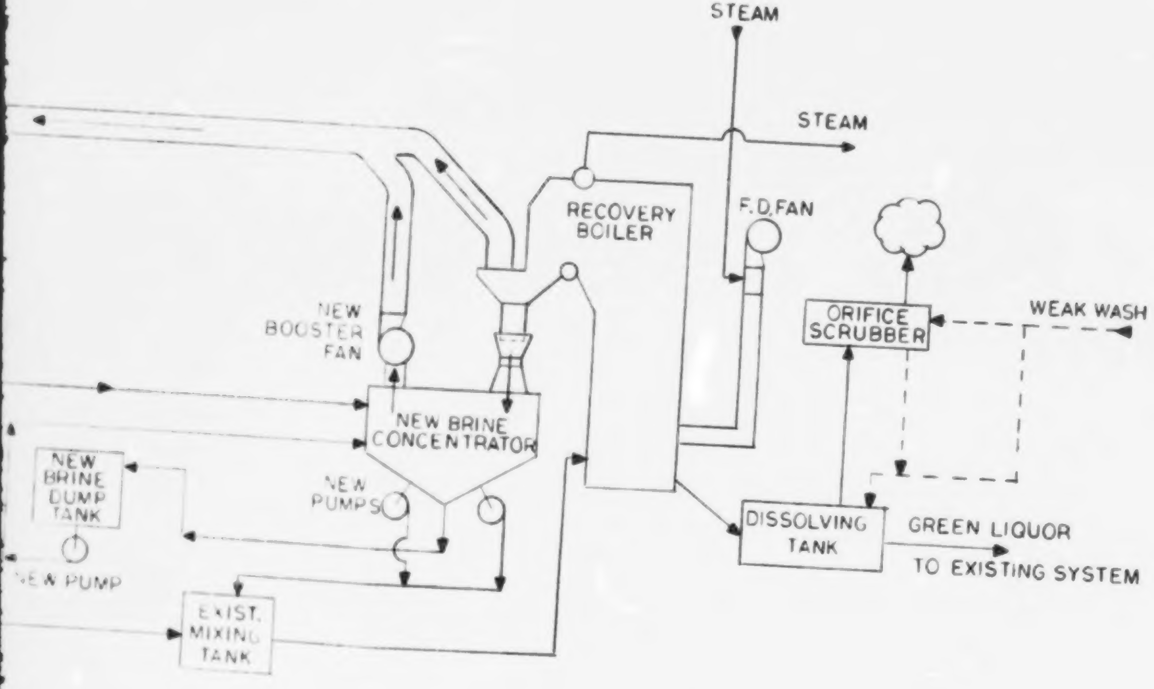
STEAM
STEAM
RECOVERY BOILER
F.D. FAN
NEW BOOSTER FAN
ORIFICE SCRUBBER
WEAK WASH
NEW BRINE CONCENTRATOR
NEW BRINE DUMP TANK
NEW PUMPS
DISSOLVING TANK
GREEN LIQUOR
TO EXISTING SYSTEM
NEW PUMP
EXIST. MIXING TANK

62% BLACK LIQUOR SOLIDS

SCHEMATIC ARRANGEMENT

CONVERT EXISTING VENTURI TO BRINE VENTURI WITH 62% BLACK LIQUOR SOLIDS

HJS

EXHIBIT NO.

FIG. 6-7

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

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GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.

Effectiveness

Particulate. Particulate removal for the above methods is estimated as follows:

a. High Solids Adjusted for 50% Collection in Cyclone Evaporator (Without Cyclone Evaporator 99.5% AOE) - 99.8% AOE

b. Wet Precipitator - 99.8% AOE

c. Na_2SO_4 Brine System - Approx. 98.0% AOE

Sulfur Dioxide. The difference in SO_2 emission for these three methods is unknown at this time. Based on operation at rated capacity in similar furnaces, the SO_2 emission will probably be approximately the same for all three methods.

Total Reduced Sulfur. All three methods should be approximately the same, except for the TRS emissions from the oxidation system of method (b). No data are available at this time to better quantify the TRS emission.

Operation. All of these systems are relatively unproven at the present time. Therefore, any discussion of operating problems would be hypothetical.

Summary

Based upon the above three control methods, the wet precipitator control method (b) indicates the best effectiveness with minimum cost.

Comparison with Chapter 5. A comparison of the wet precipitator (b) with the two stage Venturi scrubber from Chapter 5 is now in order. From paragraph 5.3.6.1.4 of Chapter 5 the two stage Venturi was an effective control method and had the lowest net annual cost. In order to achieve a TRS emission comparable to the wet precipitator control method (b), a concentrated black liquor oxidation system must be installed.

A cost comparison of these two control methods is as follows:
(500 ton size)

Control Method	Capital Cost	Net Annual Cost
Wet Precipitator	$ 1,000,000	$ 150,000
*Second Stage Venturi	$ 470,000	$ 90,000

These two control methods are considered equally effective for SO_2 and TRS removal. The particulate removal effectiveness is as follows:

Wet Precipitator	99.8% AOE
Second Stage Venturi	98.0% AOE

Considering the difference in the above costs, the second stage Venturi appears to have considerable merit if the 98.0 percent AOE particulate collection will meet the local air pollution regulations.

Therefore, for recovery systems which include Venturi evaporator/scrubbers, the addition of a second Venturi scrubber in combination with a concentrated oxidation system appears to be the most effective control method for this case. As mentioned previously, the effectiveness of the concentrated oxidation system applied to a Venturi evaporator/scrubber needs to be further evaluated.

*Broken down as follows:

	Capital Cost	Net Annual Cost
Second Stage Venturi Scrubber (from Chapter 5)	$ 270,000	$ 15,000 (from Fig. 5-34)
Oxidation	$ 200,000	$ 75,000 (from Fig. 5-21)
	$ 470,000	$ 90,000

6.3.1.4 Recovery System, Case 4

Application

This control method involves the use of molecular oxygen to oxidize weak black liquor. The method is classified as a new development since there is only one system which is currently under construction. This system has been described in detail by Galeano and Amsden (5). Oxygen can be introduced into a pipeline across a pump.

Molecular oxidation can apparently be used with either weak black liquor or concentrated black liquor. The usual problem of foaming when oxidizing 100 percent pine weak black liquor should not exist because all of the oxygen is absorbed.

The oxidation products were reported to be stable up to 12 hours. Therefore, it appears that the reversion of sodium thiosulfate to sodium sulfide, as reported by Martin (6) may not occur when using molecular oxidation. If this is, indeed, proven true in the future, this control method would probably find more application in the oxidation of weak black liquor, thereby, reducing the odorous emissions from the multiple-effect evaporators.

Costs

The capital cost for this method is rather low, and was reported by Galeano to be approximately \$75 per ADT. Therefore, the capital cost for a 500-ton mill would be approximately \$37,000.

The net annual cost would vary widely depending upon the cost of oxygen. Galeano used an oxygen cost of \$8.50 per ton, which resulted in an operating cost of \$0.07 per ADT of pulp. However, this is rather low, and is based on being located close to a pipeline supply of oxygen. From \$8.50 per ton, oxygen cost will range upward to \$100 per ton. Therefore, it is impossible to calculate any meaningful net annual cost due to the wide range of oxygen costs.

CAPITAL COST - \$37,000   NET ANNUAL COST - Wide Fluctuation

Effectiveness

Particulate. This control method has no influence on particulate removal.

Sulfur Dioxide and Total Reduced Sulfur Removal. The reduction of SO_2 and TRS would be similar to the oxidation systems described in Chapter 5.

Operation. The operation of this control method should be simpler and require less maintenance and operator attention than the oxidation systems in Chapter 5.

Summary

In summary, molecular oxidation should be equally as effective as the oxidation systems described in Chapter 5. Where the cost of oxygen would result in a net annual cost equal to the net annual costs of the systems in Chapter 5, molecular oxygen would possibly be preferred due to ease of operation and maintenance. Of course, this should be analyzed based on any additional performance data which may be available in the future.

6.3.1.5 Oxidation System

Application

The application of weak black liquor oxidation and concentrated black liquor oxidation systems to recovery units has been described in more detail previously in Chapter 5. This section discusses the application of control methods to the off-gas vent from the oxidation systems.

Data which have been published in relation to an analysis of oxidation system off-gas pollutants is very limited. Therefore, it is impossible to define the range of pollutants from oxidation systems. However, a general appraisal of some facts can be stated.

First, the off-gas flow from a weak black liquor oxidation system will usually be higher than a concentrated system. Based on a sodium sulfide loading of 10 grams/liter, a weak liquor system might require approximately 30,000 SCF/ADT of air; whereas, a concentrated system might be approximately 15,000 SCF/ADT of air. Therefore, the weak liquor system might have an off-gas flow as high as approximately double that of the concentrated system. This, in turn, means the control equipment must also be larger.

While data analyzing the off-gas pollutants are limited, Amberg and Walther (8) pointed out that a concentrated system would emit less total reduced sulfur compounds than a weak liquor system. This is attributed to the fact that the quantity of total reduced sulfur compounds in the liquor is less for concentrated liquor since a significant quantity of these compounds is vented from the multiple effect evaporators.

Costs

Costs have not been estimated because of the unknown concentrations and composition of the off-gas pollutants. Control methods which might be considered are as follows:

a. Introduction into recovery F.D. fan for incineration in recovery furnace,

b. Incineration in lime kiln,

c. Individual thermal incinerator,

d. Packed tower scrubber which uses caustic and chlorine,

e. Dilution in stack with other boilers.

Methods (a) and (b) are expected to find less acceptance because of: Increased corrosion problems in F.D. fan ductwork and F. D. fan itself; and the oxidation off-gas volume is usually too high to be handled by an existing kiln.

The remaining three methods should be investigated further.

Effectiveness

Since the concentration and quantity of pollutants are unknown, the effectiveness cannot be precisely defined. Additional research and development is needed to determine the most effective control method for the off-gas from oxidation systems.

6.3.1.6 Smelt Dissolving Tanks

Two control methods which have been discussed as possible applications to the dissolving tank are evaluated in general terms. In view of the questionable nature of these control methods, an analysis of cost and effectiveness has not been prepared. The two control methods are as follows:

Exhaust Vent Into Recovery Flue Ahead of ID Fan. For some new recovery installations, an arrangement has been discussed where the dissolving tank vent gases are exhausted into the recovery boiler flue gas stream ahead of the induced draft fan. The chemicals in the vent gases are then collected in the recovery precipitator. Since the dissolving tank vent stack diameter is approximately 3 feet to 7 feet, the routing of this stack (or duct) is difficult. For an existing installation, the rerouting of this stack is not possible without considerable expense. Due to the high capital cost, this control method is not practical for an existing installation.

For a new installation where the routing of this stack may be included in the original plans, this may be a practical control method. The most convenient method of routing the stack would be from the dissolving tank (located at the gas outlet side of the boiler) into the ductwork upstream of the cascade evaporator. However, there is some concern about the chemicals in the dissolving tank vent causing lignin precipitation in the cascade evaporator. Therefore, injection of the dissolving tank vent gases downstream of the cascade may be the most appropriate location. This location may also cause problems since the stack must be routed around the cascade to reach the downstream side. Injecting the gases into the breeching ahead of the ID fan is a potential source of damage to the fan and breeching during dissolving tank puffs and explosions.

In consideration of the relatively small capital cost difference between this control method and the installation of a scrubber or packed tower, the use of this control method is somewhat questionable. Also, the aspects of damage from puffs and the possibility of lignin precipitation, further detract from this control method. One possible benefit of this control method is better dispersion of the vent gases. This is attributed to the fact that the recovery stack is usually higher than a separate smelt dissolving tank vent stack.

One mill is currently installing this system. Operating experience from this mill should help to further evaluate this method.

At the present time, this control method would have to be classified as questionable.

Incinerate Vent in Recovery Furnace. The dissolving tank vent flow will vary depending upon factors such as: shatter jet steam pressure, stack dimensions, and amount of air leakage. In general, this flow will approximate 10 percent of the recovery boiler flue gas flow.

Incineration of the dissolving tank vent gases in the recovery furnace can be considered as a method for odor incineration plus recovery of the chemicals in the vent. The chemicals would then be recovered by the boiler precipitator or Venturi scrubber. Injection of the gases into the furnace would replace other control methods such as a scrubber or demister. The vent stack must still extend above the boiler roof to relieve dissolving tank puffs.

The injection of dissolving tank vent gases into the furnace is considered to have more disadvantages than advantages. Some of these disadvantages are:

1. The recovery furnace would be more subject to damage from dissolving tank puffs.
2. The boiler induced draft fan must handle the additional 10 percent vent flow, thus possibly reducing the combustion air available to the furnace.
3. The ductwork and fan for injection of the vent flow must be fabricated of stainless steel. The vent stack itself would be carbon steel construction.
4. Corrosion problems could be expected where the duct enters the furnace since this would be a horizontal duct and some of the chemicals would settle out. This means that corrosion of the boiler pressure parts may occur.

About the only advantage to this method, as compared to a scrubber (see Chapter 5), is the possibility of reduced capital cost.

After weighing the above factors and in consideration of other control methods, this control method was considered to be undesirable.

6.3.1.7 Digester Relief and Blow, Multiple Effect Evaporators

Application

Supplementing Chapter 5, an additional control method which might be considered is a separate thermal oxidation unit for these vents. While thermal oxidation units are used successfully in many applications, caution should be used for this source since these gases are explosive.

Therefore, this control method includes a fan for introducing air into the vent gases prior to their entrance into the thermal oxidation unit. The dilution air allows the vent gases to be handled without explosions. For batch digesters, this fan may be sized to handle the maximum flow from the blow tank without a gas accumulator. This type system has been used in Sweden (12).

Costs

The thermal oxidation unit is designed to heat the gases to 1300-1500°F for approximately 0.5 seconds. The costs as shown in Figures 6-8 and 6-9, include the necessary piping, wiring, etc., for a complete installation. The costs for the batch digester include a gas accumulator. The Swedish system which omits the gas accumulator would be somewhat less expensive.

Effectiveness

Particulate. There are no particulate emissions from this source.

Reduced Sulfur. Thermal oxidation has an effectiveness approaching 100 percent.

Sulfur Oxides Removal. The reduced sulfur compounds are oxidized to sulfur oxides; therefore, the sulfur oxides will be emitted.

Operation. Due to the physical location of these vents, the thermal oxidation units will be installed in areas which are relatively remote from operator attention. This should be considered when evaluating this control method.

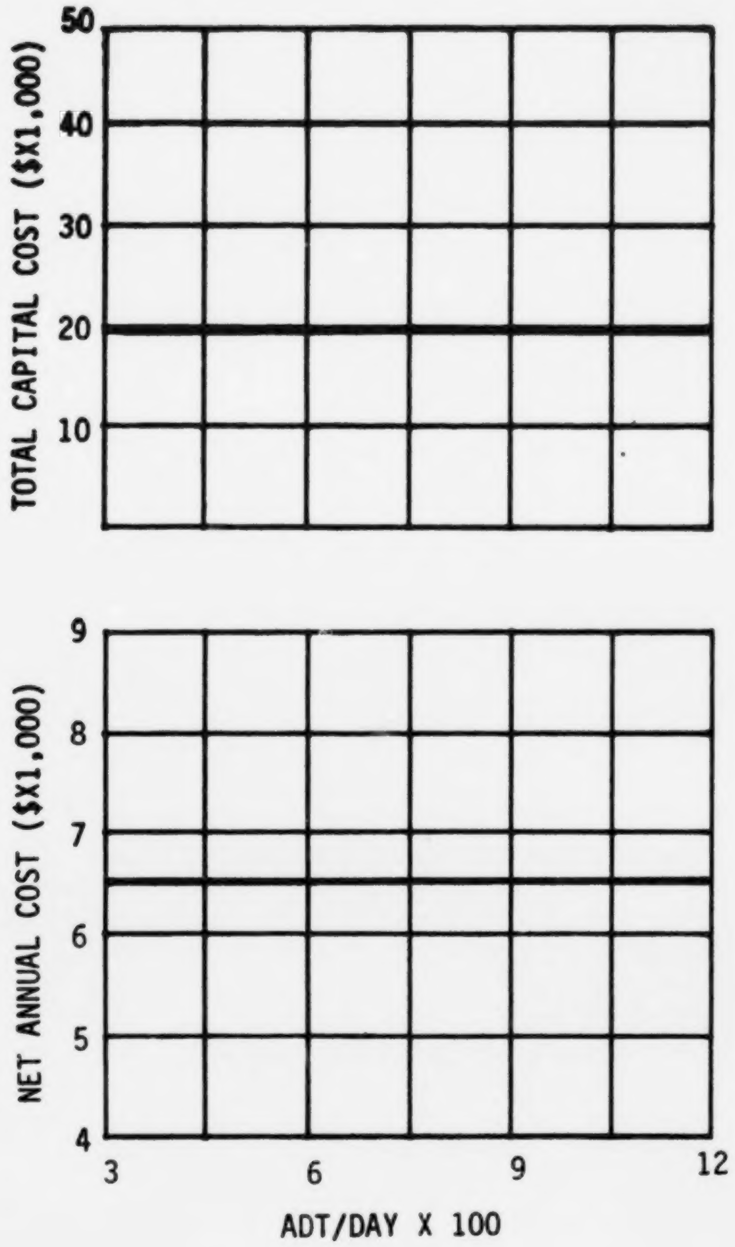


FIG.6-8 CONTROL METHOD COSTS FOR DIRECT FLAME INCINERATION OF EVAPORATOR RELIEF GASES OR CONTINUOUS DIGESTER RELIEF GASES

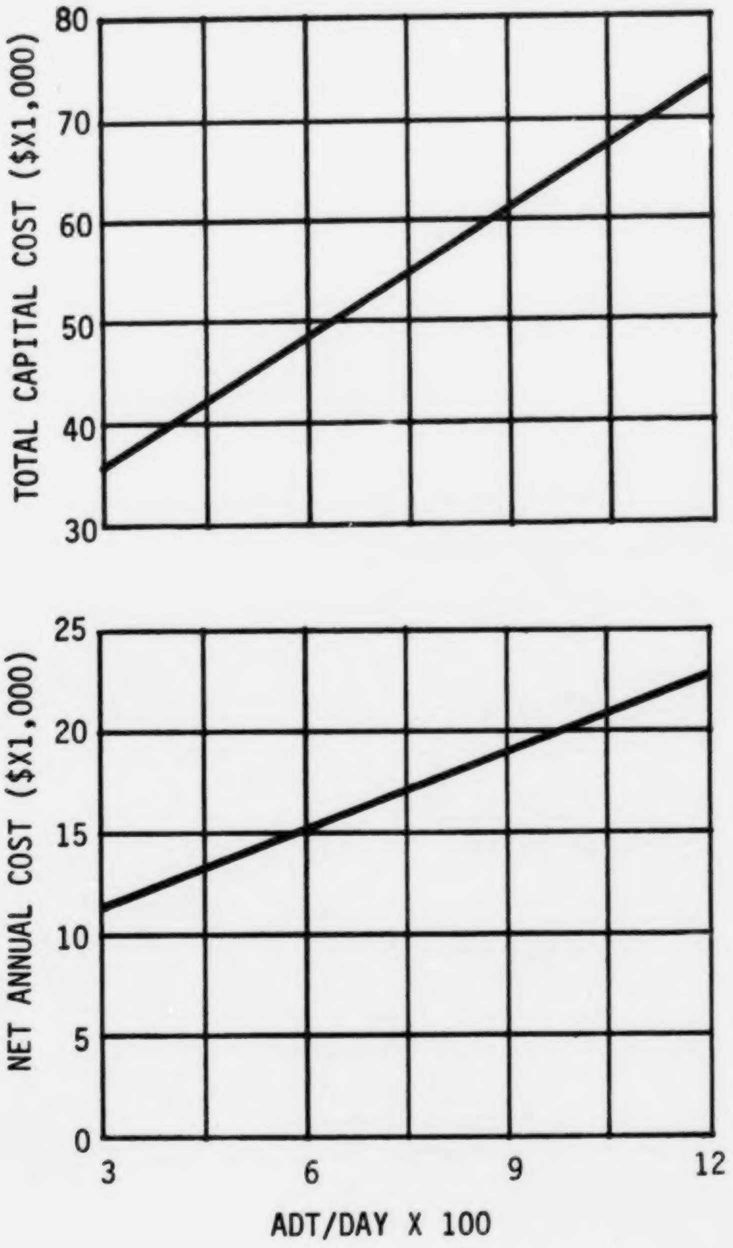


FIG. 6-9 CONTROL METHOD COSTS FOR DIRECT FLAME INCINERATION OF BATCH DIGESTER RELIEF AND BLOW GASES

Summary

The net annual cost of this control method is very similar to incineration in the lime kiln which is discussed in Chapter 5. The sulfur oxides emission from this source are higher than the control method of incineration in the lime kiln.

Therefore, the control method of incineration in the lime kiln will usually be preferred to this control method of individual thermal oxidizers. Individual thermal oxidizers will undoubtedly find acceptance in mills where the lime kilns are remote from the digesters and multiple effect evaporators.

6.3.1.8 Combination Boilers

Application

This newly developed control method combines a multi-tube cyclone collector with a shave-off scrubber. About 20 percent of the total gas flow is exhausted from the hopper of the new multi-tube collector into the shave-off scrubber. This system is used to replace existing low efficiency multi-tube cyclone collector to an overall collection efficiency of 96 percent.

For purposes of this report the following items were assumed for a system evaluation: boiler energy input of 40 percent bark and 60 percent oil, collectors and ducts of mild steel, field erection from shop assembled knock-down components, and reuse of existing sand classifiers. The material collected in the shave-off scrubber were assumed to be discharged to the existing ash handling system. The dust collector has a pressure drop of 2.5 inches WG and the scrubber requires an additional 6.0 inches WG for 20 percent of the total gas flow.

Costs

Cost curves have been based on the system gas handling capacity, since there is no direct relation between gas flow and mill tonnage. These curves are presented in Figure 6-10.

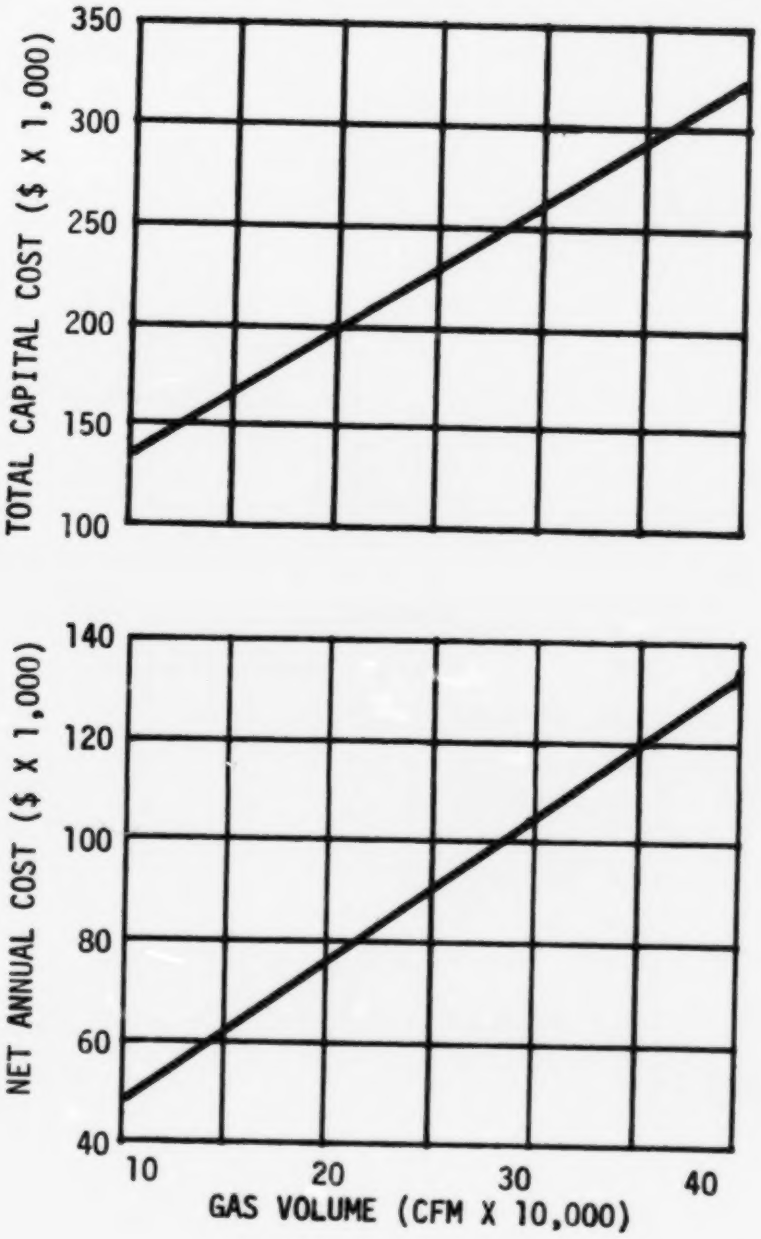


FIG.6 - 10 CONTROL METHOD COSTS FOR REPLACING MULTI-TUBE COLLECTOR WITH NEW MULTI-TUBE COLLECTOR AND SHAVE-OFF SCRUBBER - COMBINATION BOILER

Effectiveness

Particulate. This method has an expected guaranteed efficiency of 96 percent, which is equal to two multi-tube collectors in series as described in Chapter 5.

Sulfur Dioxide and TRS Removal. This control method does not remove and SO_2 or reduced sulfur compounds.

Operation. The operation of the multi-tube collector of this control method will be the same as any other multi-tube collector. However, the operation of the shave-off scrubber is not clearly defined. One consideration that may be of concern is the handling and disposal of the bark char from the scrubber. A strainer has been included for this control method; however, the handling and disposal of the scrubber effluent must be investigated for each individual application. This control method may create a water pollution problem.

Summary

For new installations, two multi-tube collectors in series (Chapter 5) will have a net annual cost approximately equal to this method. Since the two collectors in series are a more tried and proven design, the two collector methods will probably find more acceptance than the collector and scrubber method for new installations.

It is possible that the application of the shave-off scrubber may find more acceptance on existing installations where space is limited. Of course, this would have to be investigated for each application.

6.3.1.9 Brown Stock Washer Vents

The following three systems have been investigated to handle the high volume, low concentration gases from this source:

a. Continuous Diffusion Washing

This is a patented system in which pulp flows axially upward through the unit, while wash water diffuses radially. The wash water, distributed by nozzles, diffuses through the stock and is extracted through a strainer. The clean washed stock is discharged with the help of a scraper at the top of the unit. Because the unit is enclosed, no malodors should escape to the atmosphere.

Order of Magnitude Capital Cost:

500 ADT/Day	-	$ 450,000

The above price is for a single stage of diffusion washing which follows four hours of diffusion washing in a continuous digester. Prices include equipment and its erection, building, equipment foundations, instrumentation, piping and valves, filtrate tank and pump, motors, motor controls, and wiring. Indirect costs of 30 percent are included, consisting of contingency, engineering, general construction overhead, spare parts, and sales tax.

b. Enclosed Pressure Washing

The design of the pressure washer differs from the conventional vacuum filter in that the pulp mat is formed and the washing liquor is displaced from the mat by a difference of positive pressure instead of vacuum. The washer is completely enclosed with an airtight hood. An external blower provides pressure under the hood, and the air pressure forms the mat on the drum wire and passes into the drum interior. The air and vapors then discharge from the drum interior at the take-off, lifting the pulp mat off the drum. The vapors then pass to the suction of the blower to be recirculated back into the drum.

Order of Magnitude Capital Cost:

Batch Digesters. For 500 tons per day, following batch digesters and assuming three (3) washers in series, the installed cost will be approximately $1,700,000. The cost includes equipment and its erection, building, equipment foundations, instrumentation, piping and valves, filtrate tanks and pumps, motors, motor controls and wiring. Indirect costs of 30 percent are included.

Continuous Digesters. For 500 tons per day, following a continuous digester having four hours of diffusion washing and assuming one (1) washer, the installed cost will be approximately $650,000. The cost will include those items listed for washing after batch digesters and contain indirect costs of 30 percent.

c. Incineration of Vent Gases

A control method which has been suggested is the incineration of the brown stock washer vent in the recovery furnace (9). Flow from the brown stock washer vent would be introduced into the suction of the recovery forced draft fan. Since the forced draft air requirement for a recovery system is approximately 200,000 SDCF/ADT of pulp, the brown stock washer vent flow of approximately 90,000-150,000 ACF at 120° F/ADT of pulp can be accommodated by the recovery system.

The brown stock washer vent gases are usually very corrosive and will require special consideration in the selection of forced draft fan and duct materials.

For discussion, if a brown stock washer vent flow of 100,000 ACF/ADT is assumed then the following duct sizes result based on 2000 FPM velocity:

	Mill Size ADT/Day		
	300	500	1000
Approx. Flow - ACFM	21,000	35,000	70,000
Approx. Duct Diam.	3.7	4.7	6.7

In considering the application of this control method to an existing installation, the problem of material changes to prevent corrosion and the problem of physically installing a large duct in an existing facility, require that each installation would have to be evaluated individually.

This control method could be more easily applied to a new mill or a major expansion at an existing mill.

The method is a new development which has never been installed.

Summary

At present none of the three developments has been applied within the United States. There is one installation of diffusion washing of brown stock in Canada and all three methods have had limited application in Europe.

Of the three methods the practice of incineration in the recovery furnace will probably cause the most concern among operators because of possible resultant upsets in furnace operation and corrosion problems.

There are several installations of pressure washers in Europe and one in Japan. In theory, because the vapors are being recirculated within the washer, this system can reduce emissions to the atmosphere. The gas flow, if any, would be very small and, therefore, can be incinerated in the lime kiln.

Brown stock diffusion washing systems have been installed in Europe and Canada. Because the pulp is washed in complete absence of air there should be no emissions to the atmosphere.

Either pressure washing or diffusion washing will probably increase the sodium sulfide content of the liquor as compared to conventional vacuum washers. This is expected because conventional washers expose the liquor to contact with the air resulting in some oxidation of the sodium sulfide. The effect of this increased sodium sulfide concentration on the capacity of an existing oxidation system should be considered when selecting these new design washers.

Diffusion washing and pressure washing appear to have significant future potential for the reduction of emissions to the atmosphere and should be investigated further.

6.3.1.10 Multiple Effect Evaporators

Caustic Scrubbers. Caustic scrubber systems are favored on the European continent, particularly in Sweden where this system was invented.

The system consists of first a scrubber, then a condenser (heat exchanger), and finally a cyclone separator. Noncondensible gases leaving the multiple effect evaporators pass to the scrubber. The scrubbing medium may be white or weak wash liquor or a weak caustic soda solution. Ninety-nine percent of the hydrogen sulfide and some methyl mercaptan is absorbed in the caustic solution. From the scrubber the gases and liquor are discharged to a condenser, where heat is recovered. The gases and liquor then continue to a cyclone separator, the gases being exhausted through the evaporator vacuum system and the liquor returned to the causticizing system. Because the dimethyl sulfide and the dimethyl disulfide are not absorbed in the caustic solution it follows that these two gases will escape to the atmosphere.

Order of magnitude costs for an 800 ton per day mill will be $75,000.00, including erection and indirect costs. All material in contact with the liquid is stainless steel Type 304.

6.3.1.11 Miscellaneous

Steam Stripping of Combined Condensate. Liquids obtained by condensing the vapors from blow heat recovery and digester relief systems and the evaporation of the spent kraft liquors will contain steam-volatile organic compounds. These will include the sulfur-bearing compounds, later released from the mill effluent disposal system. Steam stripping the effluent condensates in a stripping column and then collecting the volatile compounds for disposal will eliminate or reduce the discharge to the atmosphere of these kraft odors. The stripped condensate could be purified to a degree that it could be returned for reuse in operations.

A typical steam stripping system will include a feed tank for storing the condensate, a heat exchanger for pre-heating the feed condensate, a fractionating or stripping column in which the feed condensate is passed counter-current to steam flow, a stripped condensate storage tank, a collection system for recovering and storing the stripped volatiles, and pumps, valves and instrumentation.

The sulfur bearing compounds that are stripped from the condensate consist of hydrogen sulfide, methyl mercaptan, dimethyl sulfide and dimethyl disulfide. In addition, some methanol and turpentine are volatized. Because the boiling points of dimethyl sulfide and dimethyl disulfide are approximately 100 to 120°C, the stripped condensate from the column may contain small amounts of these two compounds.

6.3.2 SULFITE SOURCES

6.3.2.1 Ammonium Liquor Incineration

Application

Incineration of ammonium liquor is practiced in relatively few mills; however, it is anticipated that more mills will utilize incineration in the future. This will occur as a result of efforts to decrease stream pollution which may result from the discharge of waste ammonium liquor.

Currently, there are two known acid sulfite mills (7) which incinerate liquor and both of these incinerate in combination boilers burning pulverized coal. There are additional ammonium NSSC mills and possibly some ammonium bisulfite mills, which incinerate ammonium liquor but this control method is confined to an acid sulfite mill. Two cases were considered as follows:

a. Incineration in Combination Boiler

Since the basis for the consideration of this control method is incineration of the ammonium liquor in a combination boiler, it appears reasonable to consider the recovery of SO_2 from this unit by scrubbing the flue gases with an ammonium hydroxide solution. However, there are several disadvantages which preclude the application of this method. The major disadvantages are:

Ash. Ash from the burning coal and bark will require high efficiency dust collection equipment to prevent the injection of ash into the cooking acid.

Higher Gas Volume. Recovery of SO_2 would require the absorption system to be designed for the flue gas volume leaving the combination boiler. While this volume will vary for each particular installation, it will be substantially higher than the gas volume produced from the incineration of the ammonium liquor only. An order of magnitude would be approximately twice as much gas volume for a combination boiler as compared to ammonium liquor incineration only. Therefore, the absorption system would have to handle this higher gas volume and to remove the SO_2 in a more dilute concentration.

High Excess Air. The combustion of fuels such as coal and bark usually requires a large amount of excess air (20 percent to 40 percent). As excess air increases, the formation of SO_3 in relation to SO_2 also increases. This in turn will produce higher quantities of sulfates in the scrubbing liquid, and increase the amount of ammonia make-up required for the system. The formation of sulfates with the consequent increase in ammonia make-up could be minimized by using a gas cooling tower and discharging the effluent to waste, however this would increase the waste treatment cost. Further, this cooling water would absorb SO_2, thus reducing the SO_2 available for recovery.

For the above reasons, the recovery of SO_2 from a combination boiler by scrubbing with an ammonium solution is not considered to be a practical method.

b. Incineration in Separate Boiler

To avoid the technical difficulties of SO_2 recovery from combination boiler flue gases, a separate ammonium liquor boiler and flue gas SO_2 recovery system were considered. Estimates were prepared for a 200 ADT per day ammonium (acid sulfite) mill. A full scale system utilizing ammonium liquor incineration and recovery of SO_2 is not in operation anywhere in the world so far as the contractor knows. Therefore, this control method must be considered unproven at the present time in so far as the exact SO_2 recovery efficiency is concerned.

Since it was assumed the liquor was previously fired in the combination boiler, no credit or debit has been allowed for the recovery boiler steam.

In designing the sulfur dioxide recovery system it is necessary to know the desired SO_2 concentration which may be emitted from the stack and the SO_2 concentration at the entrance to the recovery system. Both of these concentrations will vary from mill to mill, depending upon the quantity of sulfur in the ammonium liquor, stack height, stack exit velocity, meteorological conditions, topography, and other physical conditions peculiar to the individual mill. However, it appears desirable to design the SO_2 recovery system for an efficiency in the order of 95 percent. Based upon pilot plant work by Palmrose and Hull (11), an

SO_2 recovery efficiency of approximately 95 percent when incinerating ammonium liquor appears to be possible. The AOE which might be expected is unknown and must await the development of actual installations and test data from these installations.

The sulfur dioxide recovery system for this control method consists of four components: a complete boiler installation, a gas cooling tower, an SO_2 absorption tower, and a water cooling tower. The absorption of SO_2 by the cooling water and the gas cooling tower will result in a low pH liquid. Since this low pH liquid may damage a conventional water cooling tower, a stainless steel heat exchanger has been included as shown in Figure 6-11.

The required amount of makeup water is introduced into the SO_2 absorption tower to minimize the loss of NH_3 from the system and control the concentration of the combined SO_2 in the acid.

This control method does not consider special applications such as mills receiving salt waterborne logs. For these specialized cases, individual control methods and cost calculations must be prepared for each case.

(200 Ton Mill)

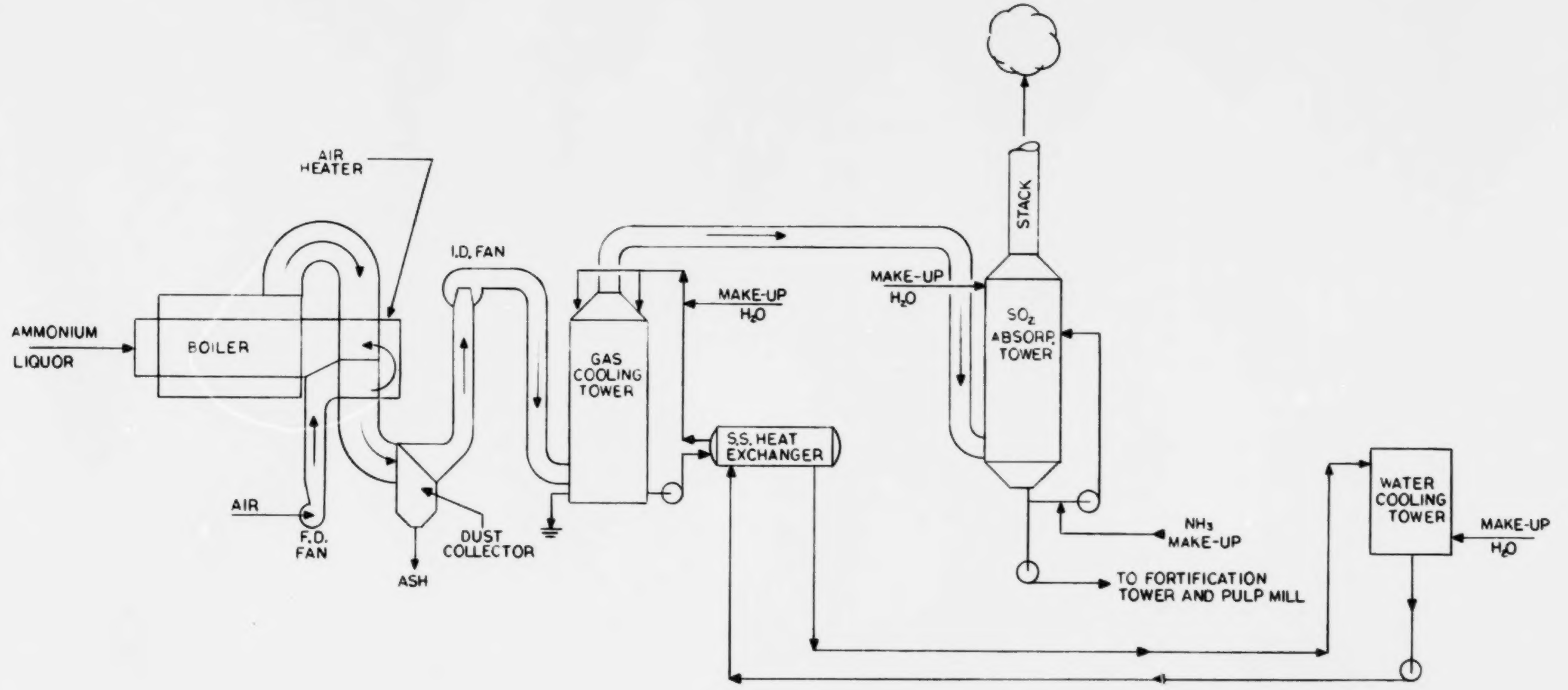
Costs

Capital Costs	$ 2,500,000
Net Annual Cost	$ 270,000

6.3.2.2 Improved SO_2 Recovery in Magnesium Sulfite Process

Application

A potential modification of the magnesium sulfite process is a reduction of SO_2 emission by upgrading the efficiency of the SO_2 recovery system. The basic process considered for this evaluation includes an MgO recovery boiler and an existing 95 percent efficiency SO_2 recovery system. As reported by Kleinegger (10), an SO_2 removal efficiency


AIR HEATER
I.D. FAN
AMMONIUM LIQUOR
BOILER
AIR
F.D. FAN
ASH
DUST COLLECTOR
GAS COOLING TOWER
MAKE-UP H_2O
S.S. HEAT EXCHANGER
STACK
MAKE-UP H_2O
SO_2 ABSORP. TOWER
NH_3 MAKE-UP
TO FORTIFICATION TOWER AND PULP MILL
WATER COOLING TOWER
MAKE-UP H_2O

H.J.S.

NEW DEVELOPMENT
AMMONIUM LIQUOR INCINERATION
AND RECOVERY OF SO_2

EXHIBIT NO.
FIG. 6-11

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

of approximately 95-97 percent may be expected from an existing installation. (Recent work indicates that these SO_2 levels may be in question.) This efficiency may be increased by up-grading the existing SO_2 recovery system by the addition of a packed absorption tower which may increase the overall SO_2 removal efficiency above 95 percent. The SO_2 concentration entering the recovery system and the SO_2 concentration leaving the stack will vary for the same reasons previously stated under paragraph 6.2.2.1 b.

The cost for this control method was developed for a 200 ADT per day mill. It was assumed that adequate space was available for the installation of this additional packed tower.

Costs (200 Ton Mill)

Capital costs include a packed tower and a booster I.D fan with the necessary pumps, piping, ductwork and electrical wiring.

Capital Cost	$ 200,000
Net Annual Cost	50,000

6.4 NSSC SOURCES

Due to the lack of emission data (as described in Chapter 5), control methods for NSSC Sources were not analyzed.

REFERENCES

1. Hochmuth, F. A., "An Odor Control System for Chemical Recovery Units," Pulp and Paper Magazine of Canada, 70, 57-66, (1969).

2. Groce, A. B., Jr. and Harris, A. D., Jr., "Evaluation of Kraft Liquor Concentrator Operation," presented at TAPPI Alkaline Pulping Conference, October 14-17, 1969.

3. Arhippainen, Bengt and Westerberg, E. N., "Kraft Odor Controls - Its Effect on Mill Operating Parameters and Costs," Pulp and Paper Magazine of Canada, 69, April 19, 1968.

4. Roberson, J. E., "How Does Recovery Odor Control Affect a Kraft Mill Energy Balance," Pulp and Paper, November, 1969.

5. Galeano, S. F., and Amsden, C. D., "Weak Black Liquor Oxidation With Molecular Oxygen," presented at the 62nd APCP Meeting in New York, New York, June 22, 1969.

6. Martin, F., "Secondary Oxidation Overcomes Odor from Kraft Recovery," Pulp and Paper, 43, June 1969.

7. Clement, J. C. and Sage, W. C., "Ammonium Liquor Burning and SO_2 Recovery," presented at 23rd Engineering Conference of TAPPI, November 6, 1968.

8. Amberg, H. R. and Walther, J. E., "The Status of Odor Control in the Kraft Pulp Industry," presented at National AICHE Meeting, Portland, Oregon, August 24-27, 1969.

9. Suda, Stanley, "Kraft Recovery Odor Reduction System," presented at 23rd TAPPI Alkaline Pulping Conference, Jacksonville, Florida, October 14-17, 1969.

10. Kleinegger, J. C., "Relative Absorptive Efficiency of Packed Towers in Magnesium Acid Bisulfite Production," TAPPI, 52, 1291-, July 1969.

11. Palmrose, G. V. and Hull, J. H., "Pilot Plant Recovery of Heat and Sulfur from Spent Ammonia Base Sulfite Pulping Liquor," TAPPI, 35, May 1952.

12. Tucker, W. G., Personal Communication, February 9, 1970.

CHAPTER 7

CRITICAL REVIEW OF CONTROL TECHNOLOGY

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CHAPTER 7

CRITICAL REVIEW OF CONTROL TECHNOLOGY

SUMMARY

In Chapters 5 and 6, control methods presently in use and new developments as applied to various sources were analyzed. The relative merits and specific limitations of the most effective and economical control methods are summarized in this chapter. An example also is analyzed of the ability of selected kraft process configurations to meet selected emission limitations.

Within each process, control devices are evaluated as to applicable emission sources, efficiency, flexibility, economics, reliability, and adaptability.

[illegible]

[illegible]

7.1 INTRODUCTION

Chapters 5 and 6 analyzed various control methods applied to different sources. The relative merits and specific limitations of the most effective and economical control methods are summarized in this chapter in general terms.

Control methods are frequently selected for new plants or new additions to existing plants while the new plant or addition is in the design stage. Therefore, parameters such as gas flow rates, temperature, composition, density, and viscosity, as well as the particulate concentration, size distribution, composition, shape, and density are all designer estimates. Once the facility is in operation, the real values for these variables become evident and may differ somewhat from design values.

Guaranteed efficiencies are often quoted by equipment manufacturers for particulate removal. These guarantees are based upon assumed values for the above parameters and must be adjusted based upon test conditions which may differ from those anticipated. The procedures for converting from guarantee to test conditions are often complicated and usually involve both the physical laws and empirical formulas.

As indicated in Section 5.3.3 most equipment manufacturers base their guarantee of collector efficiency on procedures specified by the Industrial Gas Cleaning Institute (IGCI). (See also Section 9.2.2.) The method is based on a determination of the average particulate concentration at the inlet and outlet of the collector. Isokinetic sampling at specified locations in the ducts for specified times is required at steady operating conditions. The particle collector is specified as a filter of any material and form which has a collection efficiency in excess of 99.0 percent for particulates of the approximate size distribution to be encountered during the test. Collectors other than filters are acceptable if it can be demonstrated that they have an efficiency exceeding 99.0 percent at test conditions.

The design of control methods for the removal of gaseous pollutants is less precise than the control methods for particulate removal. In actual practice, it is usually the exception rather than the rule that guarantees are made for gaseous pollutant removal.

Since changes to control methods to meet guarantees are usually costly and interrupt plant operation, control methods must be judiciously specified for both guaranteed performance and flexibility in design and operation to satisfy possible future process changes. When pollutant removal guarantees cannot be obtained from equipment suppliers or designers, the plant often finds that decisions must be made to insure conservative design.

In order to meet more stringent air pollution regulations in the future, it is possible that pilot plant test data may have to be collected and extrapolated for the design of full scale control methods. Situations of the type cited in the preceding paragraphs require the cooperation and appreciation of all segments of industry, government and the public.

7.2 KRAFT PROCESS

The kraft pulping process has received the most attention in recent years regarding emissions to the atmosphere. Since the kraft pulping process produces approximately 70 percent of the total chemical pulp in the U. S. A., it is expected that the major portion of expenditures for air quality control will be required by the existing kraft mills.

From Kraft Flow Diagram No. 10 in Chapter 3, it is apparent that the emissions from a new modern kraft mill can be nominal with the possible exception of the brown stock washers. Even these emissions from the brown stock washers can be expected to be drastically reduced as additional technology and experience are developed for diffusion washing and pressure washing as described in Chapter 6.

An attempt was made to compare emissions from older kraft mills (15 + years), older mills which have been upgraded to place emphasis on emission control, new mills (designed within the past 5 years), and mills designed in 1970 with the latest tested control technology. The emissions from kraft mills vary widely depending on many factors including product being produced, type of wood being pulped, production rate, process variables, operating variables, and combination of process equipment used.

Two older mills, operated essentially as they might have been constructed more than 15 years ago, are typified by Kraft Flow Diagrams No. 1 and No. 5. Emission control equipment in the chemical recovery system was limited to an electrostatic precipitator on the recovery boiler having an 88 percent AOE and a scrubber on the lime kiln with an 80 percent AOE. A mill of similar vintage but upgraded to reduce emissions is illustrated by Kraft Flow Diagram No. 3. Emission control systems include a 92 percent AOE precipitator on the recovery boiler, a 99 percent AOE scrubber on the lime kiln, a secondary scrubber following the precipitator having an AOE of 80 percent, weak black liquor oxidation, and incineration of noncondensibles. Mills of less than five years in age are typified by Kraft Flow Diagram No. 4. Emission control systems include a 97 percent AOE precipitator on the recovery boiler, a 99 percent AOE scrubber on the lime kiln, concentrated black liquor oxidation, mesh pads in the dissolving tank vent, and a scrubber on the off gases from the multiple effect evaporators. A mill which might be designed in 1970 incorporating the latest technological developments in emission control is illustrated by Kraft Flow Diagram No. 10. Emission control systems include elimination of the direct contact between flue gases and black liquor (thus making black liquor oxidation unnecessary), a precipitator on the recovery boiler having an AOE of 99 percent, a 99 percent AOE scrubber on the lime kiln, incineration of noncondensibles, and a 95 percent AOE scrubber on the dissolving tank vent. It must be recognized that these examples were selected because descriptions were readily available to the reader. The conditions are not the same in all cases and it must not be inferred that the systems cited represent the only way to achieve the desired reductions.

With cognizance of all of the uncertainties involved, the following tabulation is presented to compare typical emissions in terms of pounds per air dry ton of unbleached pulp. Auxiliary boilers are not included in the total.

	TRS	Particulate
Old Mill Total	18-22	18-25
Recovery Boiler	12-16	10-11
Upgraded Old Mill Total	10	3
Recovery Boiler	5	2
Relatively New Mill Total	4	4
Recovery Boiler	2	3
Latest Design Mill Total	2	3
Recovery Boiler	0.1	2

Recent emission control regulations for kraft mills such as in the states of Washington and Oregon have limitations broadly tabulated as follows (pounds per air dry ton of pulp produced):

	1972	1975
TRS Recovery Furnace	2	0.5
Noncondensibles	Incinerate	
Particulates Recovery Furnace	4	4
Particulates Other Sources	1.5	1.5

It should be noted that of the examples cited, which do not represent all possibilities, only the relatively new mill and the latest design mill can meet the 1972 TRS standards and only the latest design mill can meet the 1975 TRS standards. All except the old mill can meet the particulate standards. This points out that the state of the art of particulate collection is more advanced than that of odor reduction. Note also that the proportion of TRS produced from the recovery boiler system (including the DC Evaporator) has been reduced from 65 - 70 percent of the total to about 5 percent. As the major sources of TRS are brought under control, sources formerly considered as insignificant become of importance.

The impact of standards for the TRS emissions from kraft recovery boiler systems could be dramatic. Apparently, there is a consensus that standards set for 1972 (in Washington and Oregon) can be met by the utilization of BLO or by conversion to one of the "new" design evaporator systems. Standards proposed for 1975, however, might only be met by scrapping existing process equipment and replacing with a completely new system which eliminates direct contact between the flue gas and black liquor. If BLO can be utilized effectively, or if a conversion of existing facilities can be successfully achieved, the capital costs (for a 500 TPD mill) could range from $200,000 (for BLO) to $1.5 million (for an air contact evaporator). If it is determined, however, that a new recovery boiler, per se, is required, then the capital costs could approach $8 million.

7.2.1 PRECIPITATORS

For a more general discussion of precipitators, the reader is referred to Chapter 5.

Applicable Emission Sources. Within the chemical wood pulping industry, precipitators have found extensive applications to recovery systems; however, application to other sources in the industry is very limited. Other than direct contact evaporator recovery systems, specific precipitator applications are reported for a Venturi scrubber recovery system (See page 6-6) and a bark and coal fired combination boiler. The bark and coal fired application is the first reported installation on a bark fired boiler. This unit is expected to start up in 1970. The ability of the precipitator to collect bark char is of interest to the industry. Bark char is expected to be very difficult to collect due to the high carbon content which results in a low electrical resistivity.

Precipitators have found limited application to coal fired boilers in the industry. This is attributed to the fact that most of the coal fired units are stoker fired and their emissions heretofore have not been considered excessive. While the most modern coal fired units are pulverized fired, their capacity is usually below 500,000 lbs./hr. steam capacity and mechanical collectors have been considered sufficient control. It is expected that future coal fired units (both stoker and pulverized) will have provision for a precipitator--either initially or allowance provided for future installation. This will result from more stringent air pollution regulations.

Efficiency. Precipitators are limited to the removal of particulates only. SO_2 and TRS are not affected. Where precipitators are installed as a primary collector, their efficiency may be guaranteed as high as 99 + percent for either recovery systems or coal fired boilers. More common practice for modern design recovery systems is a precipitator efficiency of approximately 99 percent. More important than the extremely high efficiency is the reliability of operation which one buys with a modern unit of this type.

For coal fired units provisions are usually made for possible future installation. It must be recognized that regulations requiring the burning of low sulfur coal may have a deleterious effect on precipitator efficiency by altering the condition of the flue gases.

Flexibility. Precipitators are very flexible in handling load variations provided the design capacity is not exceeded. Even when the design capacity is exceeded, the precipitator will still remove particulate at a reduced efficiency.

For recovery systems, the flue gas must be at a temperature no less than 275°F - 300°F. At lower temperatures, excessive corrosion may occur in the precipitator.

For coal fired boilers a low sulfur content of the fuel may significantly reduce the efficiency of a precipitator.

Economics. Precipitators are expensive control methods (See Chapter 5). However, they are economically justified on recovery boilers up to a total annual incremental cost which does not exceed the value of the annual collected salt cake ($30 per ton has been used in this report for the cost of salt cake). In the analysis performed in Chapter 13, it will be seen that this break-even point is in the range of 90 - 95 percent for a new modern installation.

Capital costs and net annual costs, for secondary precipitators or replacement precipitators to upgrade particulate collection efficiency for an existing installation, generally range as follows per ADT/Day:

Capital Cost	Net Annual Cost
$1000 to $2000	$300 to $600

Reliability. Modern precipitator design for recovery systems utilize double or triple chamber units, multiple electrical fields, and sophisticated solid state controls. All of these provide high reliability.

Even with high reliability, unforeseen circumstances may still occur that result in high short term emissions. For example, a transformer rectifier failure may result in approximately a 10 percent decrease or more in precipitator efficiency until it can be replaced.

Adaptability. While precipitators are adaptable to a variety of applications throughout the U. S. A., their application in the wood pulping industry is primarily limited to recovery systems because of cost, corrosion, and process considerations.

7.2.2 VENTURI SCRUBBERS

Applicable Emission Sources. Venturi scrubbers have been applied to recovery systems and lime kilns in the pulp and paper industry.

For modern design, single stage Venturi scrubbers are used almost exclusively for lime kilns.

Single stage and two stage Venturi scrubbers have been used for recovery systems, with the vast majority of units being single stage. Approximately 30 recovery systems utilize single stage Venturi scrubbers while there are approximately three two stage installations. Due to relatively low particulate collection efficiency and plume visibility, Venturi scrubbers are seldom installed today for kraft mill recovery furnace applications.

Efficiency. For lime kilns, Venturi scrubber units designed for as high as 99 + percent collection of lime solids and 90 - 95 percent collection of soda fume are frequently applied for modern installations.

For recovery systems, single stage Venturi scrubber units with design efficiencies up to 94 percent particulate collection are in operation. Several two stage Venturi units have been installed for design efficiencies in the order of 99 percent particulate collection; however, test data from these installations have not been published.

Design efficiencies may vary considerably from the AOE (Annual Operating Efficiencies) of an installation. This is attributed to the fact that the collection efficiency of a Venturi scrubber is dependent upon the pressure drop across the scrubber. This in turn is dependent upon operating the unit reasonably close to design flow conditions, or providing an adjustable Venturi throat. Since existing recovery installations do not incorporate variable Venturi throat designs, the AOE may vary as much as 10 percent from the design efficiency; depending upon the operating variables at a particular mill.

The low flue gas temperature and high moisture content of the flue gases from a Venturi scrubber may be a disadvantage because of plume rise and plume visibility considerations.

Venturi scrubbers may also be effective in reducing gaseous emissions; however, this has yet to be clearly demonstrated in the industry. For lime kilns, the scrubbing medium may affect the odorous emission. Therefore, fresh water rather than contaminated condensate should be used to minimize odorous emissions due to stripping.

For recovery systems with Venturi scrubbers which utilize black liquor for a scrubbing medium, odorous gases may be either absorbed or released depending upon factors such as: concentrations of malodorous gases in the flue gases entering the scrubber, pH of the black liquor, and the sodium sulfide concentration in the black liquor. In view of the current emphasis on minimizing pollutant emissions from recovery systems, these Venturi scrubbers which utilize black liquor are not expected to receive significant consideration in modern practice.

Flexibility. Venturi scrubbers which incorporate variable throat designs are very flexible providing the design capacity is not exceeded. Flows above design capacity are usually limited due to horsepower limitation of the induced draft fan.

Economics. Venturi scrubbers are one of the lowest cost control methods, however, some specific installations may require expensive materials of construction to resist corrosion and thereby negate some of this low cost. Capital cost and net annual cost ranges per ADT/Day are:

	Capital Cost	Net Annual Cost
On Recovery Furnaces	\$300 to \$600	\$130 to \$170
On Lime Kilns	\$150 to \$250	\$ 45 to \$ 80

Reliability. The reliability of the Venturi scrubber is usually very high with most outages being attributable to auxiliary equipment such as the induced draft fan.

Adaptability. Venturi scrubbers are adaptable to collection of any particulate which may be handled in a wet system. Further, they may be adapted to the control of gaseous emissions. An example of this is the use of caustic as a scrubbing medium as discussed in Section 6.3.1.10.

7.2.3 CYCLONIC SCRUBBERS

Applicable Emission Sources. Cyclonic scrubbers are most effective as a secondary control method added to an existing particulate control device. Methods considered in Chapter 5 include the addition of a cyclonic scrubber following a recovery boiler precipitator and following a bark boiler mechanical dust collector.

Efficiency. The efficiency of cyclonic scrubbers is a maximum of approximately 85 percent on relatively coarse particles and drops off considerably for particles less than 2 microns. Cyclonic scrubbers may also be effective as gas absorbers, depending upon the particular application and the scrubbing liquid.

Flexibility. Scrubber performance will generally be affected by variations from design pressure drop. In addition, variations of particle sizes may have a pronounced effect on scrubber performance.

Economics. Costs for cyclonic scrubbers are generally quite low (See Chapter 5) as tabulated below per ADT/Day:

	Capital Cost	Net Annual Cost
On Recovery Furnaces	\$300 to \$600	\$65 to \$90
On Slakers	\$ 30 to \$ 60	\$10 to \$17

Reliability. Cyclonic scrubbers have high reliability with little possibility of plugging.

Adaptability. The adaptability of cyclonic scrubbers is generally good; however, on large capacity units, size of the equipment may become an important consideration.

7.2.4 RECOVERY SYSTEMS WITHOUT DIRECT CONTACT EVAPORATORS

Applicable Emission Sources. Recovery systems which eliminate direct contact between the flue gas and black liquor may be applied to existing recovery systems, or may be substituted for existing recovery systems. These are systems which receive black liquor at a concentration of 62 percent solids or recovery systems which utilize the air contact evaporator.

These systems minimize the emission of odorous gases but have no effect upon the particulate matter. In fact, the particulate concentration in the flue gas will increase because the direct contact evaporator, which has a capability of removing approximately 50 percent of the entering dust loading, has been omitted. Therefore, the particulate dust loading at the precipitator inlet for a system which eliminates direct contact between the flue gas and black liquor will be approximately double the dust loading for a conventional recovery system.

Efficiency. The efficiency of these new recovery systems is difficult to define in percentage terms. Currently, both of the new recovery systems marketed in the U. S. are being designed for low hydrogen sulfide emissions.

As stated previously, the particulate loading for these systems is expected to increase by a factor of approximately two at the precipitator inlet.

Flexibility. Since recovery systems which eliminate the direct contact of flue gas with black liquor are relatively new in the U. S., and the first one or two installations have been started up only late in 1969, their flexibility in operation is yet to be conclusively demonstrated and documented. However, the new recovery systems should approximate the operating flexibility of the conventional recovery systems.

Economics. The economics of these systems are more completely discussed in Chapters 5 and 6; however, of all the control methods reviewed, these new recovery systems are certainly the most expensive.

Capital costs range from $9,000 to $18,000 per ADT/Day for a new recovery system as reported in Chapter 5. For conversion of an existing recovery system (See Chapter 6) to the new design, the capital costs are $1,000 to $3,000 per ADT/Day. Because of variations in chemical recovery costs from mill to mill, net annual cost ranges are not tabulated.

Reliability. Since these new recovery systems have only recently been placed in operation, the reliability has yet to be demonstrated and documented.

While these units have been designed for low H_2S emissions, upsets in operation or equipment malfunctions can certainly be expected to increase emissions on an annual basis. For example, if upsets should occur for a total aggregate hours during a year equivalent to 14 days at a concentration of 500 parts per million H_2S; then the annual average concentration of H_2S might be increased by a factor of two or more.

7.2.5 BLACK LIQUOR OXIDATION

Applicable Emission Sources. Oxidation systems are either the concentrated black liquor type (approximately 50 percent solids) or the weak black liquor type (approximately 15 percent solids). These systems are used primarily to reduce malodorous emissions from the direct contact evaporator of the conventional recovery system. The weak black liquor type will also reduce malodorous emissions from the multiple effect evaporators. There is no effect on particulate emissions.

Efficiency. Either oxidation system can be designed for the same high efficiency of oxidation. The efficiency is based on the amount of sodium sulfide which is converted to sodium thiosulfate. This in itself is not a true indicator of the level of sodium sulfide in the black liquor which is received by the direct contact evaporator. A sodium sulfide level of approximately 0.10 grams per liter is desired to minimize odorous emission from the direct contact evaporator. This level of 0.10 grams per liter is considered to be the lowest concentration that can be accurately analyzed and guaranteed.

It should be remembered that the oxidation units themselves may be a source of odorous emissions in the exhaust gases. The effectiveness of BLO should be evaluated in terms of the TRS reduction realized at the DC evaporator, minus the TRS emitted at the oxidation unit itself.

Flexibility. The concentrated black liquor oxidation system provides more flexibility than the weak black liquor oxidation system. This is due to the fact that the weak system utilized high air flow with minimum liquor storage; while the concentrated system utilized approximately 3 - 4 hours of liquor storage with less air flow. Since the weak system has less retention time, the oxidation efficiency is primarily a function of the air flow. Therefore, if higher than design sodium sulfide loadings in the weak liquor are introduced into the weak system, a higher sodium sulfide loading will leave the unit. The additional retention volume of the concentrated system provides reaction time for fluctuations in sodium sulfide loadings and the influence of sodium sulfide variations is minimized.

As shown in Section 5.2.7, the weak system is also subject to an increase in the sodium sulfide concentration as the liquor flows through the multiple effect evaporators. Since the concentrated system is installed following the multiple effect evaporators, the concentrated system is not subject to this increase in concentration of sodium sulfide.

Economics. Installed costs for oxidation systems are detailed in Chapter 5.

Capital costs and net annual costs range from \$300 to \$800 per ADT/Day and \$130 to \$260 per ADT/Day, respectively.

For mills which operate with high sulfur losses, oxidation systems may receive a significant economic credit for chemical make-up savings in the form of salt cake. For more modern mills, oxidation systems should be classified as strictly air pollution control expenditures; since the sulfur losses are usually minimal insofar as chemical savings are concerned. Even some older mills which generate substantial quantities of by-product sulfur compounds could not achieve any chemical savings with oxidation systems.

Reliability. Adequate data are not available to assess the reliability of oxidation systems. However, if both types of systems--weak and concentrated--are designed on a comparable basis, the reliability would also be comparable.

Adaptability. With proper design, either system is considered to be equally adaptable. The concentrated system, however, has more flexibility or adaptability for handling varying sodium sulfide concentrations for a specific application. For foaming liquors, only the concentrated system can be used for oxidation.

In general, oxidation systems are considered to be the most promising control method to reduce malodorous emissions from recovery systems in older mills.

7.2.6 ORIFICE SCRUBBERS

Applicable Emission Sources. The orifice scrubber is an effective means of controlling emissions from dissolving tank vents. This application has been discussed in detail in Section 5.3.6.2.

Efficiency. Orifice scrubbers have a maximum efficiency of approximately 99 plus percent on particulate. Depending on the scrubbing liquid used, orifice scrubbers may also be effective on gaseous sulfur compounds. Scrubbing liquids are discussed in considerable detail in paragraph 5.3.6.2.a.

Flexibility. Operation of the orifice scrubber is closely related to the pressure drop (approximately 8 inches), and variations from design may have a significant effect on efficiency.

Economics. Costs have been developed in Chapter 5 and are moderate for most dissolving tank installations.

Capital costs and net annual costs range from $60 to $120 per ADT/Day and $35 to $50 per ADT/Day, respectively.

Reliability. The reliability of the orifice scrubber is usually very high with most outages attributable to auxiliary pumps and fans.

Adaptability. Orifice scrubbers may be used for other sources where moderate pressure drop control methods are acceptable. The orifice scrubber is adaptable to most dissolving tank vent configurations.

7.2.7 MECHANICAL COLLECTORS

Applicable Emission Sources. Mechanical collectors are effective control methods for boilers burning bark alone or in combination with other fuels. Bark char is difficult to collect, and the application of mechanical collectors to this material has been discussed in detail in Chapter 5. Existing installations normally utilize large diameter cyclones or multi-tube collectors arranged for single or double stage collection. Several new designs for mechanical collectors for bark have been proposed during the last few years. One such design is the addition of a shave-off scrubber as discussed in Chapter 6.

Efficiency. The effectiveness of mechanical collectors is limited to the collection of particulate with no effect on gaseous emissions. Efficiency for bark collection ranges from a maximum of 92 percent for single stage units to a maximum of 96 percent for a two stage collector or a single stage with shave-off.

Flexibility. Mechanical collectors are sensitive to particle size and density, and fluctuations in either can be expected to affect collector performance. In addition, variations from design pressure drop may significantly affect performance.

Economics. Costs for mechanical collectors have been developed in Chapters 5 and 6. These costs are usually low for normal installations. As demonstrated by the flow diagrams in Chapter 3, the required capacity of these units varies widely. Ranges of costs per ADT/Day are as follows:

	Capital Costs	Net Annual Costs
Single Stage	\$100 - \$350	\$25 - \$ 85
Two Stage	\$200 - \$750	\$50 - \$175

The cost of these devices is entirely for air quality control since there are no credits.

Reliability. The reliability of mechanical collectors can be adversely affected by plugging and hopper fires; however, these problems can generally be precluded with proper design.

Adaptability. With many arrangements available, the adaptability of mechanical collectors is very good.

7.2.8 INCINERATION IN LIME KILN

Applicable Emission Sources. Incineration of noncondensible, malodorous gases in lime kilns is a control method which has been developed by mills on the West Coast. This control method is usually applied to noncondensible gases from the multiple effect evaporators, the digester relief and blow, and the turpentine condenser.

All three of these sources have a relatively low total gas volume (in the order of 200 to 300 cubic feet/ADT). Existing lime kilns have the capability of handling these volumes. Other sources might be considered for this control method; however the quantity of gases which may be handled by an existing kiln is fixed due to the physical limitations of the installed equipment. Therefore, any additional sources must be small in volume and should be judiciously investigated.

Efficiency. Incineration does not remove any particulate matter. The malodorous gases of H_2S and TRS are almost completely destroyed by incineration. Thermal incineration converts these malodorous gases to SO_2 and water vapor. Limited field studies by industry engineers indicate that the concentration of SO_2 is removed in the lime kiln system.

Flexibility. If the incineration system is properly designed, adequate flexibility is provided for day-in, day-out operations. Of course, during malfunctions or shutdowns of the lime kiln, these gases must be vented to the atmosphere or a second auxiliary control method must be provided.

Economics. More detailed costs for this control method are provided in Chapter 5. The cost for this control method is entirely an air quality control cost since credits for recovered chemicals are insignificant. Capital cost and net annual cost per ADT/Day generally range as follows:

	Capital Cost	Net Annual Cost
Batch Digesters	\$100 - \$200	\$40 - \$75
Continuous Digesters	\$ 50 - \$100	\$20 - \$35

Reliability. In the past, this control method has only been moderately reliable. However, due to engineering refinements and more experience from those mills where it is practiced, this control method may now be considered as a reasonably tried and proven method which will provide adequate control of gaseous emissions from the above sources.

Adaptability. As previously stated, this control method is limited to relatively small gas volumes.

7.2.9 ENCLOSED PRESSURE WASHING

Applicable Emission Sources. Enclosed pressure washing is a patented system based on operating the washer cylinders with pressure rather than under a vacuum. This system eliminates the need for suspended hoods over the washers and the high exhaust gas flows.

Efficiency. An efficiency cannot be applied to pressure washers because this system would be used only as an alternative to a system using vacuum washers. Because the pressure washer is completely enclosed with an airtight hood, emissions due to washing can be practically eliminated.

Flexibility. The flexibility of pressure washing has not been documented and is unknown.

Economics. The cost of a pressure washing system is approximately the same as for a vacuum washing system. As discussed in Chapter 6, the capital cost is in the order of $1,000 to $3,000 per ADT/Day. Net annual costs were not calculated.

Reliability. Because there are no known installations of pressure washers in the U. S., it is not possible to evaluate the reliability of pressure washers.

Adaptability. Pressure washing is expected to be more applicable to a new mill installation or to an expansion at an existing mill.

For existing mills, pressure washing will probably receive more acceptance than diffusion washing; however, this must be investigated further.

7.2.10 CONTINUOUS DIFFUSION WASHING

Applicable Emission Sources. Continuous diffusion washing is a patented system designed to replace the existing system of washing with vacuum cylinders with their suspended hoods and resultant high exhaust gas flows.

Efficiency. An efficiency cannot be applied to diffusion washing because this system would be used only as an alternative to a vacuum cylinder washing system. Because the diffusion washer is enclosed and because only washed stock, relatively free of malodors, is exposed to air at the top discharge, emissions due to washing can be practically eliminated. Should washing not be complete because of overload, the small quantities of gases from the vent can possibly be piped to a lime kiln or furnace for incineration.

Flexibility. The flexibility of continuous diffusion washing has not been documented and is unknown.

Economics. The diffusion washer can be installed on the top of a high density storage tank and, therefore, there is no need for an operating floor or washer building such as required with vacuum cylinder washers. In addition, large seal tanks are not required. This results in significant capital savings. Capital cost is in the order of $1,000 per ADT/Day. Net annual costs were not calculated.

Because consistency can be kept high in the washing system, the horsepower required for predilution to vacuum washers is eliminated, creating a savings in operating cost.

Reliability. The first brown stock continuous diffuser-washer was put into operation in Sweden in 1964. The second diffuser-washer was installed in Canada in 1968 where problems were encountered due to cold weather. Minor operational upsets were encountered.

Adaptability. Continuous diffusion washing is expected to be more applicable to a new mill installation or to an expansion at an existing mill.

For existing mills, numerous factors such as space, operation, and existing equipment would probably preclude the installation of continuous diffusion washing.

7.3 SULFITE PROCESS

Control methods applied to the sulfite pulping process consist primarily of methods to remove sulfur dioxide gases. The quantity of SO_2 gases that can be released to the atmosphere from any one mill must be determined on an individual mill basis. This is a function of the ambient air quality standards, existing concentration of SO_2 in the atmosphere prior to the construction of the mill, meteorology, topography, and mill design.

Work is currently underway in some states to establish the acceptable level of emissions from a sulfite mill. Certainly, additional testing, sampling, and development work appear to be required in the field of sulfite SO_2 emission control. Reliable monitors need to be developed for these sources.

7.3.1 PACKED TOWER SCRUBBERS

Applicable Emission Sources. Packed tower scrubbers are effective control methods for reducing gaseous emissions and recovering SO_2 from sulfite blow pits. In addition, packed towers have found application to kraft dissolving tank vents.

Efficiency. The packed tower is a good gas absorber with an efficiency of approximately 95 percent on SO_2 absorption as stated in paragraph 5.3.7.2. Collection of particulate is limited because of possible dust build-up and consequent plugging.

Flexibility. Packed towers are usually very flexible, with the gaseous absorption efficiency increasing for reduced flows. However, carryover of the scrubbing medium may occur when the gas flow exceeds the design flow.

Economics. A major advantage of packed tower scrubbers is their low cost. In many sulfite applications, the value of recovered SO_2 more than offsets the equipment and operating costs resulting in a net savings (See Chapter 5).

Capital costs and net annual costs (savings) per ADT/Day range as follows:

	Capital Cost	Net Annual Cost (Savings)
Packed Tower for Acid Tower	\$ 12 - \$ 20	\$ 6 - \$ 9
Packed Tower for Blowpit	\$ 450 - \$ 750	(\$ 75) - (\$125)
Packed Tower and Condenser for Blowpit	\$1200 - \$2000	(\$300) - (\$550)

Reliability. The potential for plugging may seriously affect the reliability of packed towers when particulate is present. Therefore, packed towers must be used with care in situations where insoluble particulate is present.

Adaptability. Due to their simple design and reasonable size, packed tower scrubbers are easily adapted to most installations.

7.3.2 AMMONIUM LIQUOR INCINERATION AND SO_2 RECOVERY

Applicable Emission Sources. This control method is applicable only to ammonium base sulfite mills. This control method would probably be used when an ammonium base sulfite mill is ordered to remove the effluent from liquid waste streams and provide some other means of disposing of this waste. The SO_2 recovery system which is a part of the ammonium liquor incineration system would also be applicable to existing ammonium base mills which presently incinerate their liquor.

Efficiency. While some particulate will be removed, the quantity is not significant.

In regard to sulfur dioxide, this control method is expected to remove approximately 95 percent of the entering SO_2, as described in more detail in Chapter 6.

Flexibility. The flexibility of this control method has yet to be demonstrated and documented since it is not in operation in the U. S.

Economics. As pointed out in Chapter 6, this control method is expected to be very expensive. The capital cost and net annual cost are in the order of \$12,000 per ADT/Day and \$1,300 per ADT/Day, respectively.

Reliability. Since this control method is not in operation in the U. S., the reliability is an unknown factor.

Adaptability. Adaptability is also an unknown factor with the exception that a significant amount of space will be required for the installation.

7.3.3 IMPROVED SO_2 RECOVERY IN MAGNESIUM SULFITE PROCESS

Applicable Emission Sources. This control method involved addition of an absorption unit to an existing MgO recovery boiler with an SO_2 recovery system.

Efficiency. The concentration of particulate in this flue gas stream is present only in trace amounts. Therefore, particulate removal efficiency is not of concern. This control method would be primarily installed for additional removal of SO_2 gases. Since the installation of this control method following an existing SO_2 recovery system which operates at approximately 95 percent efficiency is not in operation in the U. S., the SO_2 removal efficiency is unknown.

Flexibility. The flexibility of this control method must also await an actual installation and testing.

Economics. Since the amount of SO_2 removal by this control method would be relatively low, the cost would almost totally be attributable to air quality control. Capital cost and net annual cost are in the order of $1,000 per ADT/Day and $250 per ADT/Day, respectively.

Reliability. Since this control method has yet to be installed, the reliability is unknown; however, the reliability should be comparable to the existing absorption units at a particular mill.

Adaptability. The adaptability of this control method for various mills will depend upon existing space conditions. While absorption towers are adaptable to other sources, this particular control method is based on the application to only one source which is an MgO recovery boiler which incorporates an SO_2 recovery system.

7.4 NSSC PROCESS

As pointed out in Chapter 5, data are not available on the pollutants from NSSC pulping processes. Therefore, the development of control methods for this pulping process must await additional data from sampling and testing.

CHAPTER 8

POWER BOILER SULFUR RECOVERY

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CHAPTER 8

POWER BOILER SULFUR RECOVERY

SUMMARY

Many mills operate on-site steam-electric generating plants in the 2 - 100 megawatt range to provide required additional energy over and above that provided by the recovery furnace. Where such power boilers are utilized, the fuels consumed are usually coal, oil, gas, or a combination of fuels. With coal and oil as the fuel, the potential for sulfur dioxide emissions is present. The combination of on-site power boilers and a process demand for soluble sulfur compounds presents a potentially unique advantage to the industry.

By mid 1970, Air Quality Regions called for by the Air Quality Act of 1967 will be designated in all of the states. Sulfur dioxide standards will be among the first to be promulgated in each state. Thus, the application of stringent sulfur dioxide emission standards to pulp mill power boilers appears inevitable. It is estimated that if the mix of fuels remains essentially as it is today, a portion of the make-up sulfur requirements could be provided from this source and theoretically all of the recovered sulfur could be used in the pulping process.

Several flue gas desulfurization processes are under active development at the present time. Six of these were considered as having potential application to this situation. Extensive technical and economic evaluation indicated that none of the processes had advanced beyond the pilot plant stage, and high capital and operating costs were predicted for application on this scale. It was concluded, therefore, that none of the processes was feasible for application to this situation in the foreseeable future.

8.1 INTRODUCTION

The complete combustion of coal and fuel oil, which contain inorganic sulfides and sulfur-containing organic compounds, results in the emission of sulfur dioxide into the atmosphere. Other oxides of sulfur may also be emitted but in quantities that are small by comparison; for example, about 40 to 80 parts of sulfur dioxide to one part of sulfur trioxide are emitted from fossil-fueled power plants.

Several techniques for the control of sulfur oxide air pollutants are under intensive investigation. These investigations have been brought about because of concern for the oxides of sulfur which are produced by such major industries as smelters and sulfuric acid producers as well as by the combustion of fossil fuels in steam-electric generating plants. The bulk of the research activity and the economic studies to date have been looked at from the point of view of the electric utility. From this vantage point it is seen that the control of sulfur oxides is achieved through concern for air quality. The material removed from the stack gases may or may not have a market value, and even if it does have value, the logistics of getting the material to a place where it can be used often makes the total scheme impractical.

In preceeding chapters of this report, heat and material balances for the pulping processes of concern were presented. It is generally agreed that the energy requirements for the entire kraft pulping and recovery systems can be supplied from the energy generated by the recovery furnace itself. If this is true, the questions logically to be asked are . . . why do pulp mills have power boilers and why consider power boiler sulfur recovery?

The answer to the first question is related to the way the industry operates. In Chapter 2 it is pointed out that most pulp is made by integrated companies and consumed captively without moving through the market place. About ten percent of the total pulp produced is, however, made by independent pulp producers without their own paper making facilities or by integrated companies producing surpluses for market.

In either case, energy is required to operate paper machines and/or to operate pulp dryers. This additional energy (over and above that required for the recovery operation) must be supplied from some external source. The usual approach is for the company to operate steam-electric generating plants (power boilers) on-site, although some mills purchase their energy.

Where power boilers are utilized, the fuels consumed are usually coal, oil, natural gas, or a combination of fuels. With coal, and oil as fuel, the potential for sulfur oxide emissions is present the same as with large steam electric generating plants. Control of sulfur oxides in this instance though can be thought of in terms of sulfur recovery. The combination of on-site power boilers and a process demand for soluble sulfur compounds presents a potentially unique advantage to the wood pulping industry.

8.1.1 LIMITS ON SULFUR DIOXIDE EMISSIONS

In January, 1969, "Air Quality Criteria for Sulfur Oxides" was released by the National Air Pollution Control Administration. Air quality criteria are an expression of the scientific knowledge of the relationship between various concentrations of pollutants in the air and their adverse effects on man and his environment. Air quality criteria are descriptive; that is, they describe the effects that have been observed to occur when the ambient air level of a pollutant has reached or exceeded a specific figure for a specific time period.

The document "Air Quality Criteria for Sulfur Oxides" indicates that "adverse health effects were noted when 24-hour average levels of sulfur dioxide exceeded 0.11 ppm for three to four days. Adverse health effects were also noted when the annual mean level of sulfur dioxide exceeded 0.04 ppm. Visibility reduction to about five miles was observed at 0.10 ppm; adverse effects on materials were observed at an annual mean of 0.12 ppm; and adverse effects on vegetation were observed at an annual mean of 0.03 ppm."

When air quality criteria are incorporated into control rules and regulations they become air quality standards.

Air quality standards are prescriptive. They prescribe pollutant exposures which a political jurisdiction determines should not be exceeded in a specified geographic area, and are used as one of several factors in designing legally enforceable emission standards. It is reasonable and prudent to conclude that, when promulgating ambient air quality standards, consideration should be given to requirements for margins of safety which would take into account long-term effects on health, vegetation, and materials occurring below the criteria levels specified above.

One can intuitively appreciate that the level of sulfur dioxide observed in the ambient air is indeed influenced by the amount of sulfur dioxide being emitted from some source or sources. It may seem then that all one would have to do is to limit the amount of sulfur dioxide being emitted to that level which keeps the ambient level below the prescribed amount. If only one source of sulfur dioxide is evident, this is a relatively straightforward exercise. There are a number diffusion equations which can be used to predict what the resulting ground level concentration will be for a given emission of sulfur dioxide. When multiple sources are considered, however, diffusion models must be developed and these suffer the indignities which only Mother Nature can explain. Then too, the nagging question always remains. . . How much of the total allowable emission shall we reserve for future sources?

Needless to say, the establishment of emission standards for sulfur dioxide is not easy. Nonetheless, emission standards have been set in some areas of the country. A typical expression for the limit of sulfur dioxide emissions is to wit: it shall be unlawful for any person to cause or permit the emission of sulfur compounds, calculated as sulfur dioxide, of more than 2,000 ppm.

Another approach to setting limits on sulfur dioxide emissions (especially directed toward fossil fuel fired combustion units) is to set a limit on the sulfur content of the fuel itself. The rationale for this approach being that it is easier to analyze for sulfur in the fuel than for sulfur dioxide in the stack gas, and that if you cut the sulfur content of the fuel by some percentage, you will automatically cut the sulfur dioxide concentration by an equivalent percentage. A corollary to this might be to provide that the concentration of sulfur dioxide in the stack gas must not exceed that which would be obtained by using a fuel with a specified sulfur content. This approach would allow a fuel with a high sulfur content to be burned but would require that at least a portion of the resulting sulfur dioxide be removed or recovered.

By mid 1970, Air Quality Control Regions called for under the Air Quality Act of 1967 will be designated in all of the states. Since the SO_2 Criteria and the Control Documents have been published, the clock is running in those Regions already designated. It will begin to run in other states when regions are designated in them. Therefore, it is inevitable that more stringent sulfur dioxide emissions standards will be imposed in many areas within the year. That these will be applied to pulp mill power boilers also would appear inevitable.

8.1.2 REQUIREMENTS ON MAKEUP SULFUR IN THE INDUSTRY

The sulfur makeup requirements, as obtained from visits to various operating mills, show a range from 22 to 32 pounds per ton of air dry pulp for the kraft industry. The sulfur makeup requirements of NSSC and sulfite processes, limited though the data are, show a reasonable correlation between the flow diagram sulfur makeup requirements and the mill visit data. Flow diagram data should not be used to estimate sulfur makeup for kraft because not all sulfur loss points are shown. The flow diagrams are oriented to atmospheric emissions of sulfur and do not have complete data for evaluation of sulfur losses in other process streams.

It is assumed that the sulfur makeup requirements for the kraft, sulfite, and NSSC processes are as follows:

1. Kraft - 26 pounds per ton A.D. pulp

2. NSSC - 60 pounds per ton A.D. pulp

3. Sulfite - without recovery - 243 pounds per ton A.D. pulp

4. Sulfite - with recovery - 51 pounds per ton A.D. pulp

These sulfur makeup requirements may be high or low for an individual mill, but are suggested as typical for the industry. These values are summarized in Table 8-1.

TABLE 8-1

SUGGESTED VALUES FOR SULFUR MAKEUP - POUNDS PER TON A.D. PULP

Kraft	NSSC	Sulfite	
	Without Recovery	Without Recovery	With Recovery
26	60	243	51

The form in which the makeup sulfur is supplied depends on the pulping process employed. Of the three pulping processes under consideration, the kraft process is the most flexible as to the form in which the sulfur can be supplied. The most common means of sulfur addition is as sodium sulfate just prior to the recovery boiler. Other forms in which sulfur may be introduced are elemental, sulfuric acid, SO_2, SO_3, sodium sulfite, and sodium sulfide. This flexibility of the form in which sulfur may be added is because of the highly alkaline nature of the black liquor which ties up any of the acid forms of sulfur as the various sodium salts and the reactions in the recovery boiler which cause the reduction of almost any sodium sulfur compound to the desired sodium sulfide.

The flexibility of sulfur addition applies only to the form of sulfur and the usual sodium-sulfur balance must be maintained to ensure the required sulfidity in the pulping liquor. Large amounts of some forms of sulfur added may require additional makeup chemicals to maintain the proper sodium-sulfur ratio. This is determined by normal mill process controls.

The NSSC pulping process requires that the makeup sulfur be added as elemental sulfur (to be burned to SO_2), SO_2 or sodium sulfite.

The sulfite process also requires the sulfur makeup as elemental sulfur (to be burned), SO_2 or the sodium, calcium, magnesium, or ammonium sulfite depending on the pulping process under consideration.

In the sulfur makeup requirements, no consideration has been given to sulfur additions which may be supplied from satellite processing such as crude tail oil production, bleaching, spent acid streams, other pulping processes in the same mill, or other operations which could conceivably supply more sulfur than the makeup requirements and thereby cause a particular pulp mill to be long on sulfur. Supplies of sulfur from such captive sources external to the pulping process can drastically alter the sulfur recovery economics. Such situations are specific to individual mills.

An example of the sulfur supplied from the processing external to the pulp mill is crude tall oil production.

Typically, about 180 pounds of black liquor skimmings (soap) are produced per ton of air dry pulp made from pine. This is the average of the range of 120 pounds to 240 pounds which is the soap yield variation due to geographical factors. The same species of pine grown in the extreme Southeast (north Florida for example) has a much higher fatty acid-rosin acid content than the trees grown more north and west. The higher fatty acid and rosin acid content yields more soap.

Normal soap requires about 295 pounds of 78 percent sulfuric acid (range 280 pounds to 310 pounds dried) per ton of soap to produce crude tall oil. The spent acid is returned to the pulp mill chemical recovery.

Assuming no sulfur losses in the tall oil processing, this would return to the mill as spent acid the equivalent of 6.8 pounds of sulfur per ton of pine pulp or about 26 percent of the 26 pounds of makeup sulfur.

It must be remembered that this factor of sulfur returned applies only to the pulp produced from pine and must be apportioned on the pine-hardwood ratio fed to the digesters.

All kraft mills have the capability of producing tall oil and any sulfur recovery projections should consider this factor. Furthermore, some mills with crude tall oil production facilities may be purchasing soap from outside sources either consistently or sporadically, which will increase the sulfur return to the mill beyond that calculated from their own soap processing. Other external sources of sulfur must be considered where applicable.

8.1.3 ESTIMATE OF FUEL USAGE AND SO_2 PROVIDED BY POWER BOILERS

In 1969, the National Council for Air and Stream Improvement in a cooperative study with NAPCA issued a steam and power boiler survey questionnaire to the pulp and paper industry. One of the questions asked dealt with the annual average consumption of coal, oil, gas, and bark. Of the questionnaires mailed out, 282 were returned.

Table 8-2 contains a summary of the fuel consumption data reported in the questionnaires. The data are stratified by use category to show the "mix" of fuels used in the industry.

The calculation for equivalent BTU for each fuel was made on the basis of heating values as follows:

Coal	14,000 BTU/Lb.
Oil	140,000 BTU/Gal.
Gas	1,000 BTU/Ft.^{3}
Bark	4,500 BTU/Lb.

TABLE 8-2

SUMMARY OF ANNUAL FUEL CONSUMPTION DATA NCASI-NAPCA STEAM & POWER BOILER SURVEY

USE CATEGORY	NO. OF REPLIES	COAL TONS	OIL BBL	10^6 FT^3	BARK TONS
Coal (Only)	65	4,163,075	--	--	--
Coal & Oil	6	443,450	717,500	--	--
Coal & Gas	3	129,850	--	3,395	--
Coal & Bark	11	1,955,800	--	--	1,181,600
Coal, Oil & Gas	6	282,100	118,333	2,414	--
Coal, Oil & Bark	10	1,589,000	2,260,200	--	840,000
Coal, Gas & Bark	4	473,900	--	4,540	153,650
Coal, Oil, Gas & Bark	4	388,850	77,933	4,442	259,350
Oil (Only)	67	--	11,509,825	--	--
Oil & Gas	23	--	1,039,950	20,657	--
Oil & Bark	16	--	7,177,400	--	1,911,350
Oil, Gas & Bark	18	--	6,310,700	74,696	3,188,325
Gas (Only)	17	--	--	22,631	--
Gas & Bark	23	--	--	75,883	2,510,550
Bark (Only)	0	--	--	--	--
No Information Available	9	--	--	--	--
TOTALS	282	9,426,025	29,211,841	208,658	10,044,825
EQUIVALENT BTU (10^{12})		263.9	17.2	208.6	90.4

BASED ON 350 DAYS PER YEAR

It is immediately obvious that data were supplied by more mills than are listed in Chapter 2. This means that some of the fuel reported in Table 8-2 was consumed at mechanical pulp mills and at paper and paperboard mills which were not on-site with a pulp mill.

Table 8-3, therefore, contains data taken from questionnaires on which a production of chemical pulp was indicated. It is seen from Table 8-3 that useable information was received from 120 mills with a combined nominal capacity of 70,263 tons per day of A.D. pulp. This represents data from 60 percent of the chemical pulp mills accounting for 64 percent of the 1968 pulp tonnage capacity.

TABLE 8-3

ANNUAL FUEL CONSUMPTION DATA - CHEMICAL PULP PRODUCERS
NCASI-NAPCA STEAM & POWER BOILER SURVEY

CATEGORY	CAPACITY TONS A.D. PULP PER DAY	NO. OF REPLIES	COAL TONS	OIL BBL	10^6 FT^3	BARK TONS
Coal (Only)	2,348	12	1,598,800	--	--	--
Coal & Oil	1,466	4	351,400	490,000	--	--
Coal & Gas	240	2	71,750	--	2,275	--
Coal & Bark	3,443	9	1,749,300	--	--	1,146,600
Coal, Oil & Gas	185	2	149,800	35,833	1,687	--
Coal, Oil & Bark	8,265	10	1,589,000	2,260,117	--	840,000
Coal, Gas & Bark	1,380	2	439,600	--	3,507	172,900
Coal, Oil, Gas & Bark	2,820	4	388,850	77,933	4,442	259,350
Oil (Only)	2,561	10	--	3,696,967	--	--
Oil & Gas	2,421	8	--	23,718	12,191	--
Oil & Bark	9,747	14	--	6,944,650	--	1,911,525
Oil, Gas & Bark	17,270	18	--	6,281,300	72,403	3,178,350
Gas (Only)	2,596	4	--	--	9,770	--
Gas & Bark	15,521	21	--	--	73,307	2,330,300
Bark Only	--	0	--	--	--	--
No Information Available	4,440	3	--	--	--	--
TOTALS	74,703	123	6,338,500	19,810,518	179,582	9,839,025
Equivalent BTU (10^{12}) Based on 350 Days Per Year			177.5	11.6	179.6	88.6

Taking the totals from Table 8-3 and scaling them upward on the basis of 1968 pulp production ratioed to 80 percent of the nominal capacity reported (86, 571/56, 210) we get an estimate of the total quantities of fuel consumed in power boilers in 1968 in locations where chemical pulp is produced on-site. This estimate is shown in Table 8-4. A figure of 80 percent of nominal capacity reported was used because a comparison of data in Table 2-1 with data in Table 2-4 suggests that on the average, mills were operating at 80 percent of capacity.

TABLE 8-4

ESTIMATED FUEL CONSUMPTION - 1968 AT MILLS WHERE CHEMICAL PULP IS PRODUCED

Coal	Oil	Gas	Bark
9,761,290 Tons	30,508,198 Bbl.	276,566 x 10^6Ft^3	15,152,099 Tons
273.3 x 10^{12}BTU	179.4 x 10^{12}BTU	276.6 x 10^{12}BTU	136.4 x 10^{12}BTU

Again, taking data from Table 8-3 one can generate Table 8-5 which represents the quantities of the various fuels consumed per year per ton of A.D. pulp per day. Please note that these figures do not necessarily equate themselves on an energy basis. They merely represent historical data that say, for example, that 113 tons of coal were consumed in 1968 for each ton of chemical pulp produced per day.

TABLE 8-5

RATIO OF FUEL CONSUMED PER YEAR PER TON OF CHEMICAL PULP PRODUCED PER DAY

Coal	Oil	Gas	Bark
113 Tons	252 Bbl.	3.2 x 10^6Ft^3	175 Tons
3.2 x 10^9BTU	2.1 x 10^9BTU	3.2 x 10^9BTU	1.6 x 10^9BTU

If the "mix" of fuels remains constant through 1980, then the data in Table 8-5 can be coupled with data in Table 2-3 to estimate fuel consumption for the years 1975 and 1980. These estimates are given in Table 8-6.

TABLE 8-6

PROJECTION OF FUEL CONSUMPTION

- CHEMICAL PULP PRODUCERS -

	Coal	Oil	Gas	Bark
1975	14,270,092 Tons	44,451,968 Bbl.	404,109 x 10^6Ft^3	22,099,700 Tons
	399.6 x 10^{12}BTU	261.4 x 10^{12}BTU	404.1 x 10^{12}BTU	198.9 x 10^{12}BTU
1980	17,563,364 Tons	54,710,656 Bbl.	497,370 x 10^6Ft^3	27,199,900 Tons
	491.8 x 10^{12}BTU	321.7 x 10^{12}BTU	497.4 x 10^{12}BTU	244.8 x 10^{12}BTU

It is now a straightforward calculation to estimate the amount of SO_2 generated by the combustion of these fuels in pulp mill power boilers. If one assumes a sulfur content of two percent for both coal and oil--this is taken as being representative of the values given on the steam and power boiler survey questionnaires--then the values in Table 8-7 can be generated.

TABLE 8-7

TONS OF SULFUR DIOXIDE PRODUCED BY POWER BOILERS

- ALL CHEMICAL PULP PRODUCERS -

Year	Coal	Oil
1968	390,452	192,202
1975	570,804	280,047
1980	702,535	344,677

Basis 2% S for coal and oil
Tons Coal x 0.02 x 2 = Tons SO_2
BBL Oil x 42 x 7.5 x 1/2,000 x .02 x 2 = Tons SO_2

8.1.4 MAKEUP SULFUR - SULFUR LOSS RATIO

The sulfur makeup requirements presented in Table 8-1 can be coupled with data in Tables 8-7 and 2-1 to give a national picture of the balance between sulfur makeup requirements and the sulfur which conceivably could be supplied from power boiler flue gases. This relationship is shown in Table 8-8.

TABLE 8-8

INDUSTRY-WIDE DATA ON MAKEUP SULFUR REQUIRED VERSUS SULFUR AVAILABLE FROM POWER BOILERS

Process	Sulfur Required - Tons
Kraft	315,900
NSSC	105,000
Sulfite Without Recovery	218,700
Sulfite With Recovery	17,800
TOTAL	657,400
Potential Sulfur Available	291,327

It is readily apparent from Table 8-8 that from the national standpoint, recovery of all sulfur lost in power boiler flue gases would still not supply the total required makeup sulfur.

The question might well be asked then, "How does this relate to specific mill circumstances?" Is it conceivable that a given mill could lose more sulfur through power boiler flue gases than is required for makeup, or does the national picture prevail even at the individual mill level?

To determine if the sulfur makeup requirements derived above are compatible with the potential sulfur available from power boiler flue gas, the following analyses were made:

1) Kraft -- Looking at the energy balance provided for each kraft flow diagram one sees that the range of fuel required is from 386 pounds/ADT to 1180 pounds/ADT. Assuming the coal to have a sulfur content of two percent, this equates to a potential sulfur loss of 7.7 pounds to 23.6 pounds/ADT.

2) NSSC -- The range of fuel required for the NSSC processes is from 1192 to 1550 pounds/ADT. Again, this equates to a potential sulfur loss of 23.8 pounds to 31 pounds/ADT.

3) Sulfite -- The range of fuel required for the various sulfite processes is from 507 pounds to 1630 pounds/ADT. This equates to a potential sulfur loss of 10.1 pounds to 32.6 pounds/ADT.

In each case presented above, the potential sulfur recovery (sulfur loss from power boilers) is less than the corresponding sulfur makeup requirement for the specified process. It is concluded then, that if regulations are passed which require the control of SO_2 emissions from power boilers, sulfur recovery could be considered because the potential amount of sulfur to be obtained could all be utilized in the pulping processes.

The examples cited above were based only on using coal as a fuel. For equivalent energy, less oil than coal is required so that the corresponding sulfur loss would also be less. Since all the sulfur available from coal burning can be utilized, it is apparent that all sulfur lost from oil burning can likewise be utilized.

8.1.5 FUELS CONSUMED - PAPER AND ALLIED PRODUCTS INDUSTRIES

As a comparative exercise, it was decided to estimate the total amount of fuel utilized by the paper and allied products industry (SIC Group 26). Contact was made with the National Coal Association, the American Petroleum Institute, and the American Gas Association regarding compilations of fuel usage data. All three organizations reported that they do not compile specific data on types of industries. The American Gas Institute did have available gas consumption data for the paper and allied products industries for selected years. All three organizations referred to the Census of Manufacturers of the U. S. Department of Commerce.

The data reported in the Census of Manufacturers are given in detail only once every five years; the latest available being the 1963 report covering fuel usages in 1962. In addition to the detailed Census of Manufacturers, the Department of Commerce publishes an Annual Survey of Manufacturers in which data are presented for total purchase price paid for fuels by industry classification, but it does not report a breakdown of cost by types of fuels.

Table 8-9 reports the total purchased cost of fuels versus the total chemical pulp production for selected years from 1954 through 1966 plus the calculated cost per ton of pulp. It must be remembered that this total cost reflects the entire paper and allied products category and not just pulp mills. Because of fluctuations in this cost, the five year average for 1962 - 1966 of $12.07/ton pulp was chosen as a basis for projection. The projected values of dollars to be spent on purchased fuels for the years 1975 and 1980 are also shown in this table. The projected values were obtained by multiplying the cost per ton figure of $12.07 by the projected tonnage of chemical pulp.

Table 8-10 was prepared showing the breakdown of this total cost into types of fuel assuming that the national distribution by types of fuels will remain unchanged from the 1962 values.

An interesting comparison can be made by viewing the estimates of Table 8-6 with the estimates in Table 8-10. This comparison reveals that of the estimated fuel consumed by the paper and allied products industry, 48 percent of the coal, 91 percent of the oil, and 87 percent of the gas is utilized by chemical pulp producers.

TABLE 8-9

TOTAL PURCHASED COST OF FUEL vs. CHEMICAL PULP PRODUCTION FOR SELECTED YEARS

YEAR	TOTAL PURCHASED FUEL COST ($1000)	CHEMICAL PULP PRODUCTION (MILLION TONS)	COST $/TON PULP	
1954	220,036	16.0	13.75	
1958	266,444	17.6	15.13	
1959	274,797	19.8	13.87	
1960	281,064	20.7	13.57	
1962	302,885	23.1	13.11	
1963	305,464	24.6	12.41	12.07
1964	311,554	27.2	11.45	
1965	334,336	28.8	11.60	
1966	355,607	30.1	11.81	
PROJECTED				
1975	555,220	46.0	12.07	
1980	679,541	56.3	12.07	

TABLE 8-10

FUEL USAGES PAPER AND ALLIED PRODUCTS

1975 and 1980

YEAR	TOTAL PURCHASE COST ($1000)	COAL		OIL		GAS		OTHER*
		COST ($1000)	QUANTITY (1000 Tons)	COST ($1000)	QUANTITY (1000 Bbl)	COST ($1000)	QUANTITY ($10^6 Ft^3$)	COST ($1000)
1975	555,220	239,855	29,215	127,145	48,715	161,569	464,279	26,651
1980	679,541	293,562	35,757	155,615	59,623	197,746	568,236	32,618

*Gasoline, LPG, Wood and Purchased Steam

BASIS: (1) Projection of total purchased cost from Table 8-8

(2) Assumption that distribution by type of fuel and the cost per unit of each type will remain unchanged from 1962

8.2 FLUE GAS DESULFURIZATION TECHNOLOGY

Many flue gas desulfurization processes have been proposed and a number of them are currently being actively developed. All of these processes are primarily directed toward the removal of sulfur oxides from flue gas for air pollution control. The recovery of sulfur in any marketable form is a secondary feature which would help defray the cost of achieving the primary objective. We are concerned only with the recoverable sulfur processes.

Of the many processes being proposed for the removal of sulfur oxides, the state of development of most must still be considered as experimental despite the fact that some are being advertised as commercial processes. Though the technology may be sound, as demonstrated by pilot plant and semi-works scale operations, full commercial size plants have not been built as yet.

8.2.1 A CRITICAL REVIEW OF APPROPRIATE PROCESSES

The following processes are considered as having a potential application for the recovery of sulfur from power boiler flue gas.

1. Catalytic oxidation
2. Beckwell Process (Wellman-Lord)
3. Stone and Webster Process
4. Reinluft Process
5. Alkaline Scrubbing
6. Kiyoura - T.I.T.

8.2.1.1 Catalytic Oxidation (2, 3)

A promising system producing sulfuric acid from the SO_2 in the flue gases involves catalytic oxidation of SO_2 to SO_3. Some research on catalytic oxidation has been done by Bituminous Coal Research, Inc., an arm of the National Coal Association, with laboratories in Monroeville, Pennsylvania, but currently attention is focused on work now being done at the prototype plant built by Monsanto Company and Metropolitan Edison Company at the latter's Portland, Pennsylvania, station along the Delaware River.

The process consists of oxidizing catalytically the sulfur dioxide found in flue gases to sulfur trioxide, which is condensed to sulfuric acid and can be used commercially.

In 1962 the Pennsylvania Electric Company, Air Preheater Company, Research Cottrell Corporation, and Monsanto participated in the construction and operation of a small pilot plant at Penelec's Seward, Pennsylvania, generating station.

The project proved the technical feasibility of the process, but a number of questions are left unanswered. Monsanto and Metropolitan Edison decided to build and operate a prototype plant at Portland to obtain economic and engineering information necessary for the design of commercial units. Construction was begun in 1966 and completed in August 1967. The prototype plant takes only a portion of the total flue gas from the generating station, but it incorporates commercial type equipment about the size used on many industrial boilers. The development program required an expenditure of between $4 and $5 million.

Monsanto believes results are so encouraging that the process can be made available commercially in the near future.

In the process, flue gases direct from the boiler are first passed through mechanical and electrostatic collectors to remove in excess of 99.9 percent of the particulates. This efficiency is required to prevent fouling of the catalyst. The gases are then passed through a converter where more than 90 percent of the SO_2 is oxidized to SO_3 with a vanadium pentoxide catalyst.

The SO_3 rich stack gas then enters a rotary heat exchanger where incoming combustion air is used to cool the gas from about 900 degrees F to 200 degrees F. During cooling the SO_3 combines and condenses with the moisture normally present in the stack gas and forms 70 percent sulfuric acid. Most of the acid is collected on the cool surfaces of the condenser. Acid mist remaining in the gas is collected in a second electrostatic precipitator. A typical effluent for this process would contain 200 ppm SO_2 by volume and 20 ppm SO_3. Figure 8-1 shows the Catalytic Oxidation SO_2 recovery process.

8.2.1.2 Beckwell (Wellman-Lord) (12, 13)

A pilot installation, owned and operated by Wellman-Lord, Inc., of Lakeland, Florida, performed satisfactorily for several months at one of the Tampa Electric Company's Gannon Plant units. It is reported to remove more than 90 percent of the sulfur oxide and all of the flyash that remains after the gases pass a precipitator.

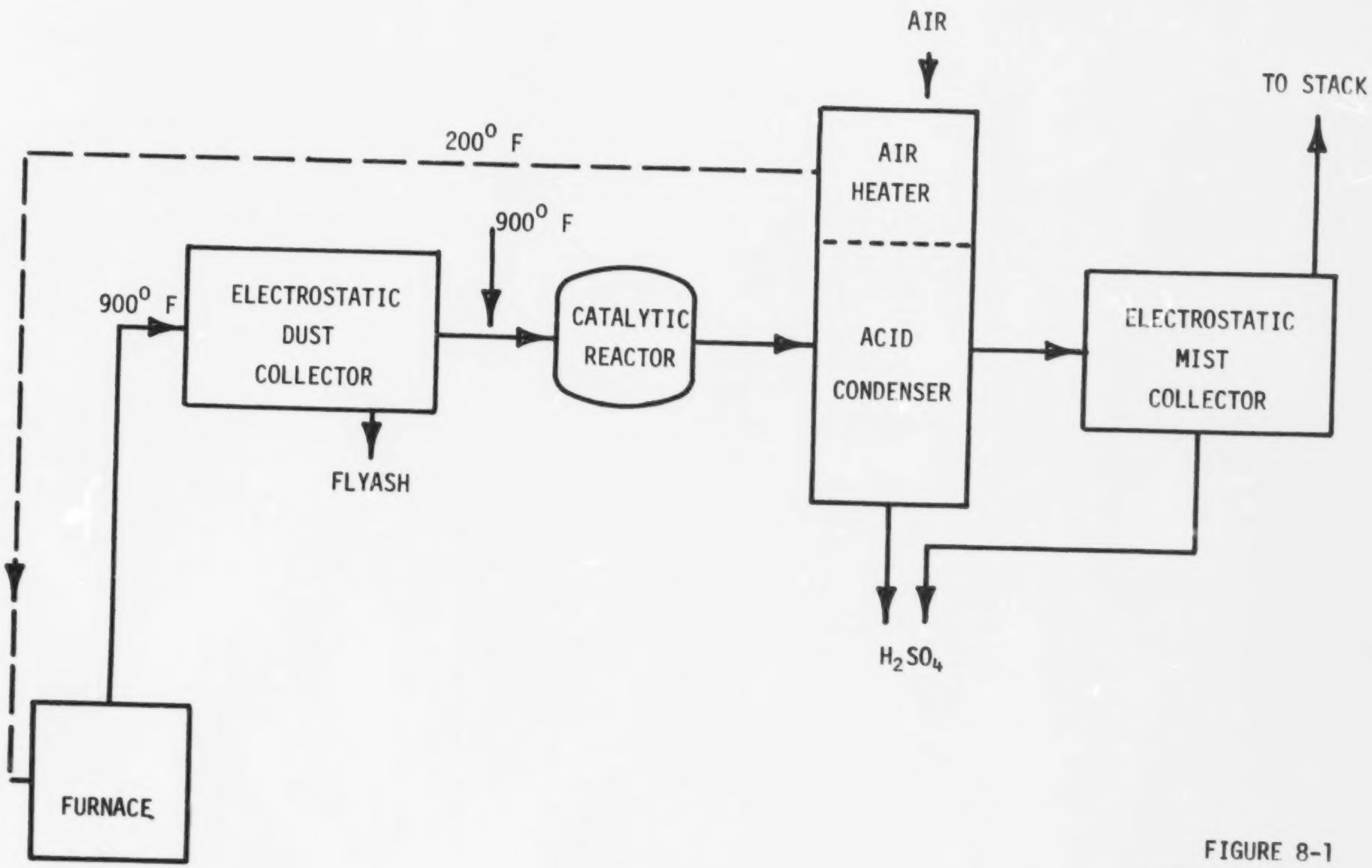
AIR
TO STACK
200° F
AIR HEATER
900° F
900° F
ELECTROSTATIC DUST COLLECTOR
CATALYTIC REACTOR
ACID CONDENSER
ELECTROSTATIC MIST COLLECTOR
FLYASH
H_2SO_4
FURNACE

FIGURE 8-1
CATALYTIC OXIDATION
SO_2 RECOVERY PROCESS

The process involves a reactor for scrubbing SO_2 from the flue gas with an aqueous solution of potassium sulfite. The SO_2 reacts to form potassium bisulfite. About 99 percent of the flyash and SO_3 are removed, along with 90 to 95 percent of the SO_2. This portion of the pilot plant operated continuously with little difficulty and minimal pressure drop. The flue gases are discharged from the reactor to the stack at 140 degrees F with a velocity of 70 to 90 feet per second. The flue gas is 70 percent saturated and plumes will not be visible until the temperature drops below 45 degrees **F**. No need for stack reheat is predicted.

The reacted scrubbing liquor is pumped to a special treating area where the potassium bisulfite is crystallized out of solution as potassium pyrosulfite, is pumped to the top of the stripping column in the recovery area, and flows countercurrent to steam injected in the bottom of the column. The potassium pyrosulfite in the heated solution reverts to potassium sulfite and releases SO_2. Approximately four pounds of steam reportedly are required per pound of SO_2 produced. More recent data indicate that this figure may be low. The potassium sulfite is recycled to the reactor scrubber.

Steam from the column is condensed and recycled from the K.O. drum to the special treatment area. The SO_2 vapor from the K.O. drum is compressed and fed to a distillation column and is condensed overhead with water. A portion of the liquid SO_2 is refluxed to the column and the remainder fed to storage.

A demonstration plant was built at the Crane Station of the Baltimore Gas and Electric Company to recover the SO_2 released from 25,000 kw of power generation. Results of this demonstration should be available in early 1970. Figure 8-2 shows the Wellman-Lord, Inc. SO_2 recovery process.

8.2.1.3 Stone and Webster Process (19, 15)

Stone and Webster Corporation and Ionics, Inc., have developed a process the basis of which is an electrolytic cell which regenerates two necessary process fluids. The process utilizes a high efficiency wet scrubber of proven design and can produce dry high purity sulfur dioxide. The process is adaptable to existing boilers. This process was piloted on flue gas from coal-fired boilers at the F. T. Gannon station of the Tampa Electric Company and is now being planned for a demonstration plant.

Further process details are unavailable at this time.

8.2.1.4 Reinluft (3, 9, 10, 15)

This process, developed in Germany, uses activated char to absorb SO_2 from the flue gases, and can produce sulfuric acid. In the process, the activated char is in a fixed, slowly moving bed. Flue gas, at a temperature above its dew point, enters the bottom of the absorber, where the SO_2 is then drawn off, cooled to 220 degrees F and returned to the absorber at a higher level.

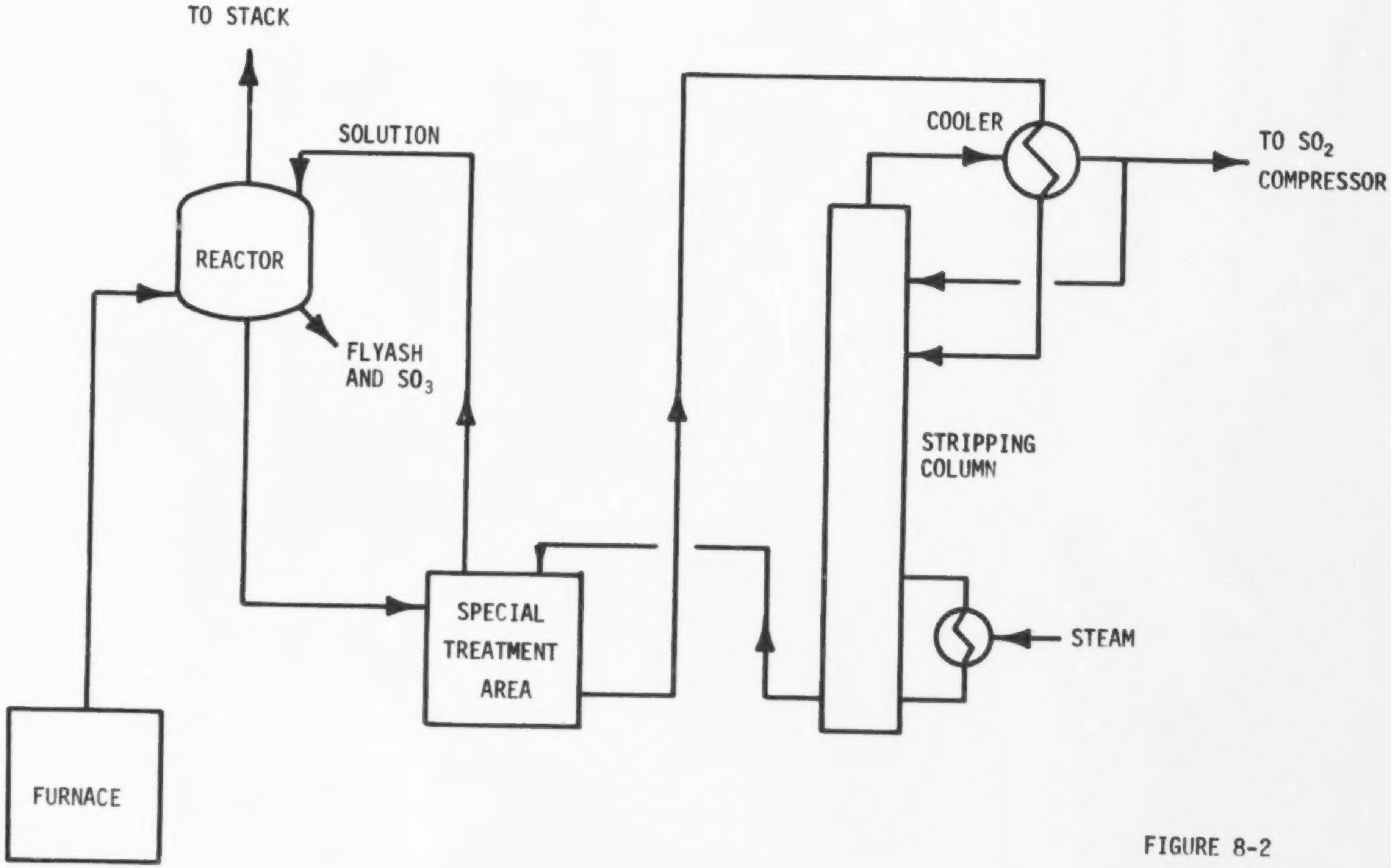
TO STACK
SOLUTION
REACTOR
FLYASH AND SO_3
COOLER
TO SO_2 COMPRESSOR
STRIPPING COLUMN
SPECIAL TREATMENT AREA
STEAM
FURNACE

FIGURE 8-2
WELLMAN-LORD
SO_2 RECOVERY PROCESS

Sulfur dioxide in the gas is oxidized to SO_3 and adsorbed as sulfuric acid along with moisture in the char. The char, with its adsorbed sulfuric acid, drops to the regenerating section where at 700 degrees F the sulfuric acid disassociates. The resulting SO_3 is reduced to SO_2. The SO_2 gas leaving the generator at 300 degrees F is heated to 700 degrees F and returned to the base of the generator. A side stream removes the SO_2 product gas.

Although the process seemed almost ready for full-scale commercial application, technical difficulties developed concerning the high consumption rate and ignition of activated char, and its applicability has become questionable. Reinluft, Inc. has recently signed over full rights to the process to Chemieban- Zieren.

One reference (16) states that three full-scale Reinluft plants either are in operation or about to be completed in West Germany. A smaller plant has been operating at the Volkswagen works for several years and Volkswagen management is installing the first of the three aforementioned plants in expectation of reducing their pickling acid purchases. Figure 8-3 shows the Reinluft, Inc. process.

8.2.1.5 Alkaline Scrubbing

A system reported by Galeano and Harding (11) uses soda ash liquor to scrub SO_2 from the power boiler flue gases, resulting in a liquid suitable for use in NSSC cooking liquor.

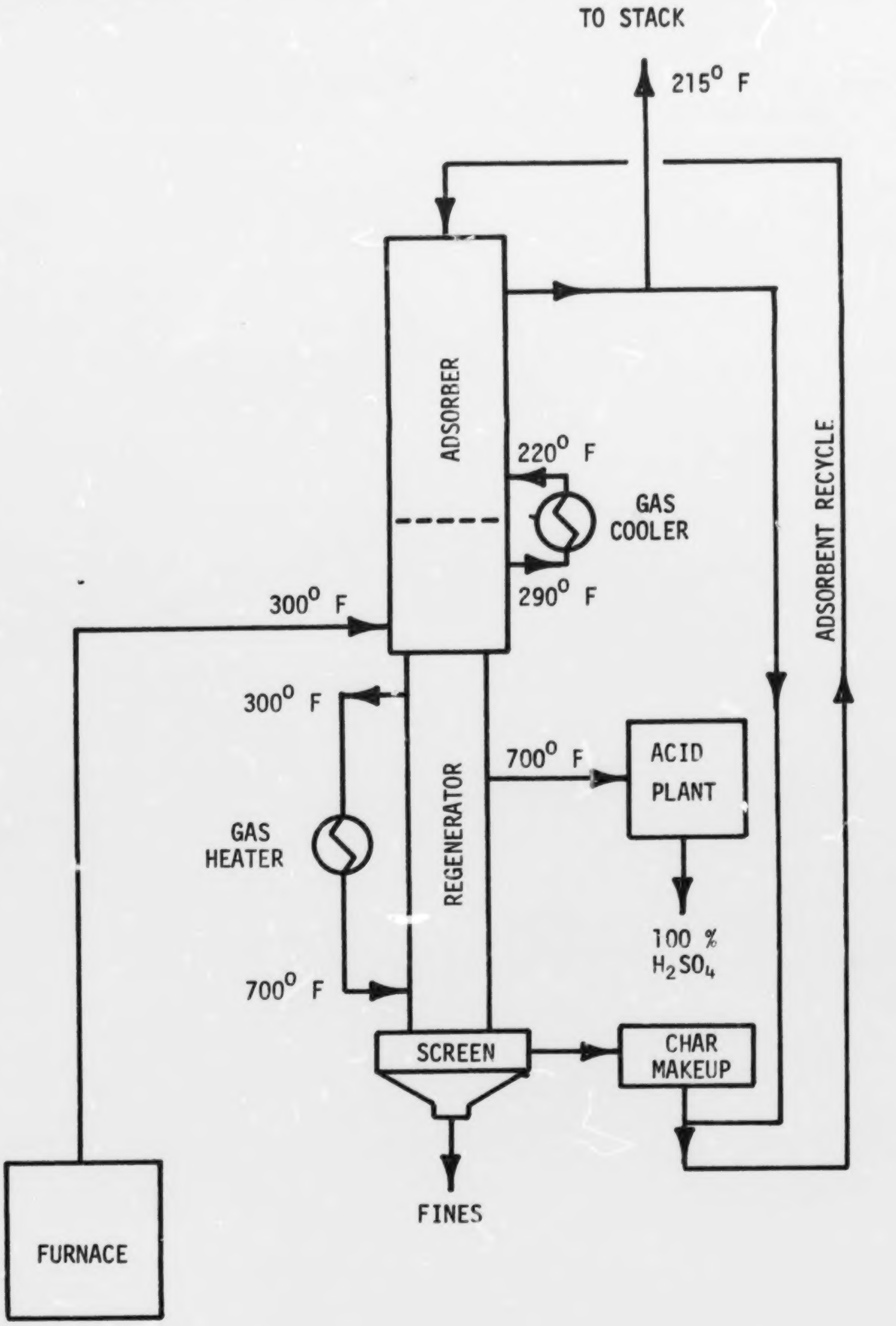
TO STACK
215° F
ADSORBER
220° F
GAS COOLER
290° F
300° F
ADSORBENT RECYCLE
300° F
REGENERATOR
700° F
ACID PLANT
GAS HEATER
100 % H_2SO_4
700° F
SCREEN
CHAR MAKEUP
FINES
FURNACE

FIGURE 8-3
REINLUFT
SO_2 RECOVERY PROCESS

The efficiency of base scrubbing for removing SO_2 from boiler stack gases generally depends upon the concentration of the base, liquid/gas ratio, temperatures, pressure drop, and the particular scrubber. Operating the system for most efficient removal of SO_2 produces a liquor volume exceeding the expected daily makeup requirements (9). To match the liquor produced to that needed, the concentration and liquid/gas ratio are the most easily changed variables.

Sulfur dioxide can be removed from a gas stream by putting it in contact with a solution of any base. Bases suitable for sulfite pulping include, $Ca(OH)_2$, NaOH, NH_4OH, and $Mg(OH)_2$. A solution of any of these can be used to effectively scrub SO_2 from the gases and produce a bisulfite solution. This weak acid can be incorporated into the acid production system of the sulfite mills, for a net savings in elemental sulfur requirements.

It is possible that alkaline scrubbing could cause some conversion of sulfite and bisulfite to to sulfate. The result of this would be the introduction of an inert impurity into the pulping liquor.

8.2.1.6 Kiyoura - T.I.T. (12)

This process uses catalytic oxidation followed by a gaseous ammonia reaction to produce ammonium sulfate. It is similar to the catalytic oxidation (Monsanto-Metropolitan Edison) method in that the hot flue gases from the boiler are cleaned of essentially all particulates before passing through a converter where the SO_2 is oxidized to SO_3 with a vanadium pentoxide catalyst.

Gaseous ammonia is added after the catalytic converter, when the stack gas, now rich in SO_3

has been cooled to 425 degrees F to 500 degrees F. At these temperatures, the SO_3 will not have condensed, and ammonium sulfate is formed. The ammonium sulfate is collected in a second electrostatic precipitator. The sulfate crystals are extremely pure and can be used directly as a fertilizer material. More than 70 percent of the flue gas SO_2 is recovered (2). Figure 8-4 shows the Kiyoura process.

TABLE 8-11

SUMMARY OF SULFUR OXIDE RECOVERY PROCESSES

Process Name	Type of Process	Form of Recovered Sulfur	State of Development	Possible Pulp Process Application	Remarks
Catalytic Oxidation	Uses V_2O_5 Cat. O Oxidize SO_2 to SO_3	H_2SO_4	Offered Commercially	Kraft	Requires high degree of particulate removal from gas (99.9 percent)
Beckwell (Wellman-Lord)	Scrub SO_2 into Pot. Sulfite Decompose Pot Pyro sulfite to rec. SO_2	SO_2	Demonstration Plant built	Kraft NSSC Sulfite	
Stone & Webster Process	Scrubbing into process fluids electrolytically regenerate solutions	SO_2	Demonstration plant planned	Kraft NSSC Sulfite	
Reinluft	Adsorb SO_2 on activated char. oxidize to SO_3 and desorb	H_2SO_4	Conflicting reports from technical difficulties to full plant operation	Kraft	Char fires common
Alkaline Scrubbing	Scrub SO_2 from flue gas with alkaline solution	Sulfite Salts	Pilot Plant	Kraft NSSC Sulfite	Use in pulping process depends on scrubbing medium chosen
Kiyoura T.I.T.	Special case of Cat. oxidation	$(NH_4)_2SO_4$	Pilot Plant	None	Produces salable $(NH_4)_2SO_4$

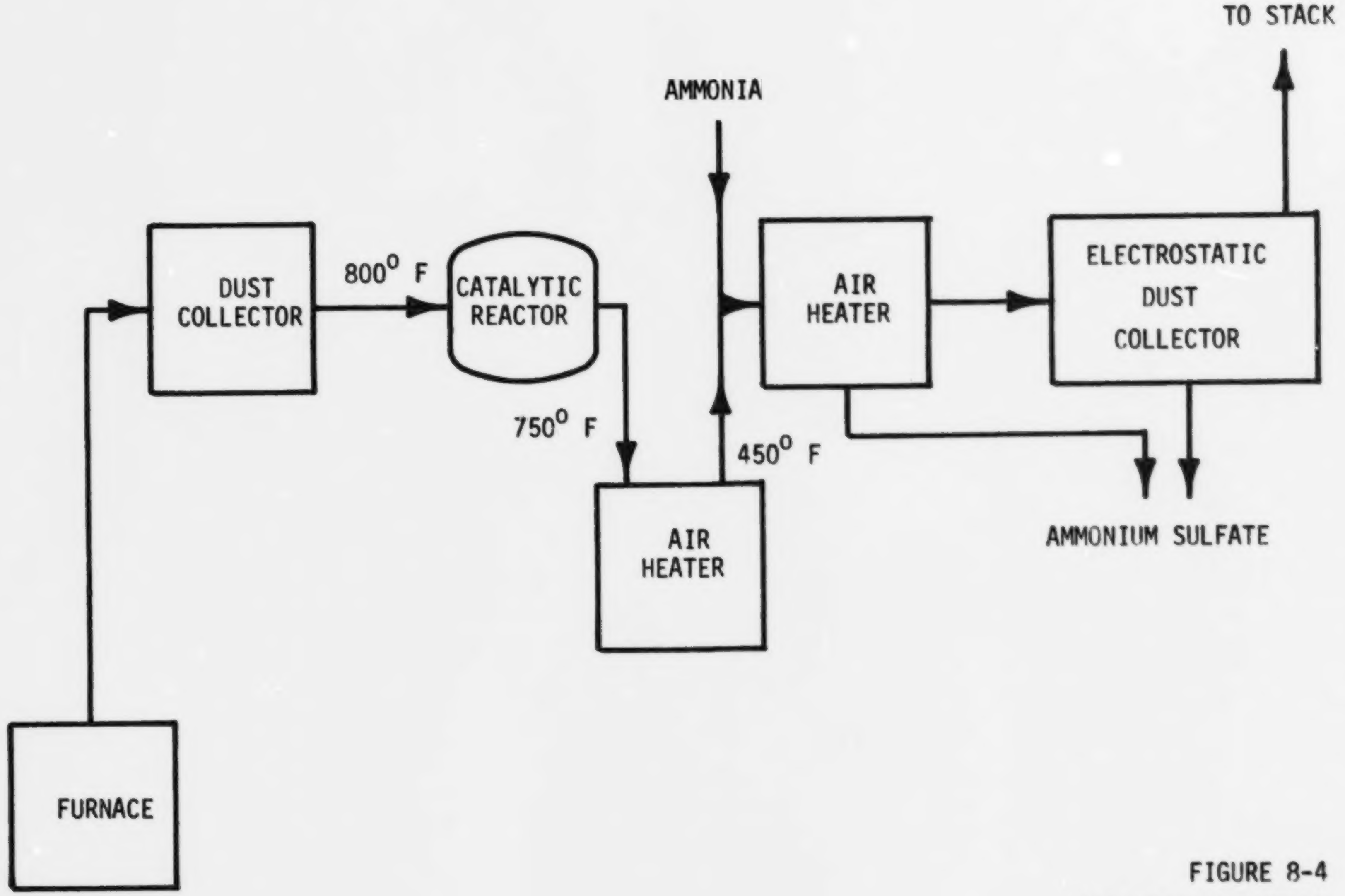
TO STACK
AMMONIA
DUST COLLECTOR
800° F
CATALYTIC REACTOR
750° F
AIR HEATER
450° F
AIR HEATER
ELECTROSTATIC DUST COLLECTOR
AMMONIUM SULFATE
FURNACE

FIGURE 8-4
CATALYTIC OXIDATION - AMMONIA
SO_2 RECOVERY PROCESS

8.3 PROCESS FEASIBILITY CONSIDERATIONS

There are many schemes, either proposed or being developed, for the removal or recovery of sulfur oxides from flue gas from utility power boilers.

Progress in developing suitable flue gas desulfurization processes has been slow because of the magnitude and complexity of the problem. Flue gas desulfurization process development is further complicated by a wide variation in the size of utility plants.

The technical and economic feasibility of most processes is closely related to plant size (15).

It is unlikely that a single flue gas desulfurization method will be developed that is capable of controlling effluents from all types of sources. The recovery technique employed will depend on factors such as boiler size and configuration, age, load pattern, fuel characteristics, and utility of by-product sulfur or sulfur compounds.

In evaluating the feasibility of the several processes outlined in order to select the most attractive, the controlling criteria are technical feasibility and economic considerations.

The major problem associated with the evaluation of the processes is the fact that most of the systems have not been developed commercially and projected cost data reported by the developers should not be taken at face value. Cost analyses based on published data seldom provide sufficient detail with which the reported economics can be verified.

Letters were written to the developers of the several processes under consideration requesting the most recent and detailed capital and operating costs that they would feel free to furnish. No replies were received from any of these inquiries therefore evaluation and comparison of the systems must be based on the limited data which appear in publications.

In addition to the lack of dependable economic data, process details necessary to reasonably estimate capital and operating costs are unavailable, presumably, because of the proprietary nature of the information. Here also, we must rely on published articles as a gauge of the technical feasibility of the systems.

Much of the published data, both economic and process, is repetitive and articles published by the developers tend to be more optimistic about the merits of their process than do the articles published by independent writers.

A recent article (17) classifies some of the various sulfur oxide recovery processes by the following criteria:

1. First Generation
    a. Advanced pilot plant studies
    b. Active research and development
    c. Adaptability to U. S. market
    d. Available data on economic assessment
2. Near First Generation
    a. Meeting first generation criteria except:
        1. Fewer advanced pilot studies or insufficient information
        2. Better adaptability to foreign operations
3. Second Generation
    a. Essentially bench studies of new concepts
    b. Pilot studies incomplete
    c. Economic assessment not practical

The considerations used in the classification were:

1. Technical feasibility
2. State of development
3. Process applicability
4. Economic factors
5. Relative advantages
6. Problem areas.

Using this system of ranking, the six processes discussed in 8.2.1 may be classified as follows:

1. First Generation
    a. Catalytic Oxidation
    b. Reinluft
2. Near First Generation
    a. Beckwell (Wellman-Lord)
    b. Kiyoura - T.I.T.
    c. Alkaline Scrubbing
3. Second Generation
    a. Stone and Webster - Ionics, Inc.

The determination of the most feasible process for the recovery of sulfur oxides from power boiler flue gases depends on the following factors:

1. Utility of recovered product
2. Technical feasibility
3. State of the development of the process
4. Economics of the process

In Section 8.2.1, the number of processes appearing attractive was found to be six, based on the criterion of recovered product utility in one or more pulping processes.

In this section, based on the criteria of technical feasibility and the state of development, the number of processes to be considered as attractive has been reduced to two which have been listed as First Generation Processes. These processes yield recovered products suitable only to the kraft pulping process.

8.3.1 Sulfur Recovery and Reuse - Economic Evaluation

The literature was reviewed in an effort to determine the parameters necessary to make an economic evaluation of the processes under consideration. All available publications appear to be based on public utility stations of 800 MW size (2.7×10^9Btu/hr) and new plant installations. Of the information gleaned from the literature sources, there is a similarity in capital and operating costs which implies that the data may have been extracted from a limited number of original estimates. This situation is created by the lack of economic data sources available to independent authors. The economic data which are available, in general, come from the developer of the particular process, and such economic data are sparse.

This study is specifically oriented to the power boilers in the pulp industry and few, if any, of these are beyond a size of 100-200 MW equivalent. (See Appendix B for distribution of boiler sizes in the pulp industry.) It becomes apparent, then, that any published capital cost will suffer in accuracy by reducing from an 800 MW to a 100-200 MW basis since the relationship of capital cost to unit size is not known. It is felt that the application of the standard "six-tenths power" factor normally used for scale-up effect on capital cost should not be used because of the uncertainties of the published capital cost data and the lack of information on the equipment details of the various processes. This is further complicated by the fact that all data have been projected from relatively small pilot plant installations. The literature gives capital costs for 800 MW installations in the range of $10 - 20 million or $10 - 22 per kilowatt. (17, 18, 19, 20, 21, 22).

Even with a reasonable reduction in capital cost attributable to scaling down from 800 to 100 - 200 MW, the investment for these processes would still be substantial and could approach approximately 10 percent of a new 800 - 900 TPD mill investment. It is felt, however, that the chances of a substantial reduction in investment are marginal because all of the investment figures shown are based on new plant installations and there are several comments in the literature pertaining to the difficulty or impracticability of putting these sulfur recovery processes into existing power plant installations due to lack of available space.

In existing installations, the generation of space sufficient to accommodate the sulfur recovery process units would insure a substantial cost which could conceivably approach the cost of the unit installation. This situation is particularly prevalent in many existing pulp mills.

In view of the above, it is obvious that under any circumstances, the installation of the sulfur recovery processes would result in a very substantial capital cost.

Operating cost data for the three processes are subject to the same limitations as are the capital cost data in that the information appears to be repetitive and probably from the same parent sources. The operating costs stated in the literature sources show a new operating cost ranging from about $0.75 to $1.50 per ton of coal and $0.40 to $0.50 per barrel of oil, including credits for the sulfur by-products recovered.

It is felt that these operating costs must be considered as minimum costs because the factors included in operating costs will not be reduced proportionately to plant size and, in fact, the operating costs may increase per unit of fuel as the plant size is reduced. Also, it should be noted that any credits included above are based on sulfur prices in the range of $35 per long ton. Recent publications (23) have indicated that this sulfur price may no longer be accurate and, in fact, may be substantially higher than future sulfur prices. Also, these credits do not include any allowances for costs to supply sodium to correct any imbalance created in the sodium-sulfur ratio.

In view of the fact that these sulfur recovery processes are not advanced beyond the pilot plant stages, and, in addition, with the high capital and operating costs predicted, it must be concluded that the processes will not be feasible for the control of sulfur oxide emissions from utility boiler flue gases in the foreseeable future.

The state of development of the non-recovery processes is no further advanced than that of the recovery processes. These processes also require additional development at the utility boiler scale before they can be applied to pulp mills.

Thus, it appears that the only feasible alternate for immediate application available for the control of sulfur oxide emissions from utility boiler flue gases for the pulp industry must be based on fuel substitution.

8.4 R & D EFFORTS

The fact that the sulfur removal processes have not been developed to a feasible stage does not preclude the interest in continuing the research and development efforts in these and similar sulfur removal processes as methods of control of sulfur oxide emissions from industrial power plant installations utilizing fossil fuels.

It is estimated that as many as eighty to one hundred separate processes or modifications are being investigated by many agencies. Currently, most of this work is being directed toward emissions control from large power generating stations (800 MW). It is suggested that these efforts should be continued and supported with additional emphasis being placed on the smaller power generating stations found in many industrial plants as auxiliary operations. In this continuing R & D program, we should not lose sight of the various fuel desulfurization programs which are being intensively studied. Some support is justified in this area as well as flue gas treatment.

Another area of needed research is to search for methods to protect the equipment used in these sulfur recovery processes against corrosion. This consideration is a continual one which affects both process and emission control equipment.

Work must also be done to devise ways to prevent process contamination by carbon when sulfur recovery processes are considered. It has been found that carbon, which is removed in particulate form from the flue gas, passes through the other process steps to show up at the paper machines. This quite obviously impairs the quality of paper being produced. Sophisticated filtration processes must be used to remove the carbon and these should be researched further.

8.5 REFERENCES

1. Libby, C. E., Pulp and Paper Science and Technology, 1, Pulp, (1962).

2. Bovier, Ralph F., "Sulfur-Smoke Removal Systems," Proceedings of the American Power Conference, 26, 135-143, (1964).

3. Moody, J. E., "Research Must Meet the Demand," Combustion, (May 1968).

4. Galeano, S. F., "Removal and Recovery of SO_2 in the Pulp Mill Industry," Dissertation, University of Florida (August 1966).

5. Harding, C. I. and Galeano, S. F., "Utilization of Weak Black Liquor for SO_x Removal and Recovery," TAPPI 50, (10), (October 1968).

6. Shah, I. S. and Stephenson, W. D., "Weak Black Liquor Oxidation System: Its Operation and Performance," TAPPI 51, (9), (September 1968).

7. Murray, F. E., "Oxidation of Kraft Black Liquor," in Atmospheric Emissions from Sulfate Pulping, (E. R. Hendrickson, Ed.), (April 1966).

8. Aires, R. S. and Newton, R. D., Chemical Engineering Cost Estimation, McGraw-Hill Book Company, Inc. (1955).

9. Kaiser, Alex, "Air Pollution Control in the Power Industry," presented at Annual Meeting, Florida Section APCA, October 7, 1968.

10. "Outside U.S., Tough Laws Spur SO_2 Removal Efforts," Chemical Engineering, 84-88, (November 4, 1968).

11. Galeano, S. F. and Harding, C. I., "Sulfur Dioxide Removal and Recovery from Pulp Mill Power Plants," Jour. APCA, 17, (8), (August 1967).

12. Terrana, J. D. and Miller, L. A., "New Process for Recovery of SO_2 from Stack Gases," presented at American Power Conference, Chicago, April 1968.

13. "Profit in Stack Gas?", Chemical Week, 53-54, (July 20, 1968).

14. Galeano, S. F. and Amsden, C. D., "Weak Black Liquor Oxidation With Molecular Oxygen," presented at 62nd Annual Meeting of the Air Pollution Control Association, New York, June 22, 1969.

15. "Control Techniques for Sulfur Oxide Air Pollutants," U. S. Department of Health, Education and Welfare, NAPCA Publication No. AP-52, (January 1969).

16. Frankenberg, T. T., "Sulfur Removal: For Air Pollution Control." Mechanical Engineering, (August, 1965).

17. Corteizon, C. G., "Commercial Processes for SO_2 Removal," Chemical Engineering Process, (September 1969).

18. Persh, E. A., Rusanowsky, N. P., Young, N. W., "An Appraisal of Air Pollution in the Power Industry," presented at American Power Conference, Chicago, April 1968.

19. "Research Goal: Solution of the Sulfur Problem," Coal Research, Bituminous Coal Research, Inc., Vol. 25, (Winter 1968).

20. Reid, W. T., "Sulfur Oxides Control in Central Station Power Plants," Heating, Piping and Air Conditioning, 153, (March 1968).

21. Kiyoura, R., "Studies on the Removal of Sulfur Dioxide from Hot Flue Gases as a Measure to Prevent Air Pollution," presented at 59th Annual Meeting of the Air Pollution Control Association, San Francisco, June 20-24, 1966.

22. Katell, S., "Removing Sulfur Dioxide from Flue Gases," Chemical Engineering Progress, 62, (10), (1966).

23. "Market Newsletter," Chemical Week, 105, (13), 43-44, (September 27, 1969).

APPENDIX B

Appendix B consists of two tables which show the distribution of steam generating capacity by on-site power boilers in the chemical wood pulping industry.

Table B-1 summarizes the total quantities of steam generated by the industry.

Table B-2 presents information on the steam generating capacity of individual power boilers in the industry.

TABLE B-1

DISTRIBUTION OF TOTAL QUANTITIES OF STEAM GENERATED BY POWER BOILERS AT 289 PULP AND/OR PAPER MILLS IN THE U.S.

Total Annual Average Steam Generation			
1,000's lbs/hr	MW Equivalent	Number of Mills	Percent of Total
0 - 20	0-1.8	24	8.5
21 - 50	1.8-4.5	48	17
51 - 100	4.5-9.0	47	16
101 - 200	9.0-18.0	53	18
201 - 500	18.0-45.0	75	26
501 - 1,000	45.0-90.0	35	12
1,000 - 1,500	90.0-135.0	7	2.5
> - 1,500	> 135.0	0	0
		289	100

Source: NCASI - NAPCA questionnaire survey of steam and power boiler in the pulp and paper industry (1969).

TABLE B-2

DISTRIBUTION OF STEAM GENERATING CAPACITIES OF INDIVIDUAL POWER BOILERS AT 282 PULP AND/OR PAPER MILLS IN THE U.S.

Steam Generating Capacity 1,000's lbs/hr	Steam Generating Capacity MW Equivalent	Number of Boilers	Percent of Total
0 - 20	0-1.8	155	16
21 - 40	1.8-3.6	187	19
41 - 60	3.6-5.4	159	16
61 - 80	5.4-7.2	119	12
81 - 100	7.2-9.0	82	8.2
101 - 150	9.0-13.5	111	11
151 - 200	13.5-18.0	67	6.8
201 - 300	18.0-27.0	66	6.7
301 - 400	27.0-36.0	20	2.0
401 - 500	36.0-45.0	15	1.5
501 - 600	45.0-54.0	7	0.7
601 - 700	54.0-63.0	0	0
701 - 800	63.0-72.0	1	0.1
		989	100.0

Source: NCASI - NAPCA questionnaire survey of steam and power boilers in the pulp and paper industry (1969)